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The objective of this seminar/workshop for consulting engineers, planners, and other professionals was to exchange information on recent research and development activities in the United States, Canada, and Europe on the design, operation, control, and testing of aeration systems.

The 3-day seminar/workshop consisted of short, formal discussions followed by an afternoon workshop session. Papers and current technical data on selected topics were presented in several technical sessions including aeration system design, operation, and maintenance; aeration system control; specification and testing of aeration equipment; and translation of oxygen transfer data from clean water to process conditions.

A summary of the afternoon workshop sessions is also presented as a portion of these proceedings.

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August 2-4, 1982

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and Environment Canada

Edited by

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OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
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## DISCLAIMER

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# FOREWORD

The U.S. Environmental Protection Agency is charged by Congress with protecting the Nation's land, air, and water systems. Under a mandate of national environmental laws, the agency strives to formulate and implement actions leading to a compatible balance between human activities and the ability of natural systems to support and nurture life. The Clean Water Act, the Safe Drinking Water Act, and the Toxics Substances Control Act are three of the major congressional laws that provide the framework for restoring and maintaining the integrity of our Nation's water, for preserving and enhancing the water we drink, and for protecting the environment from toxic substances. These laws direct the EPA to perform research to define our environmental problems, measure the impacts, and search for solutions.

The Water Engineering Research Laboratory is that component of EPA's Research and Development program concerned with preventing, treating, and managing municipal and industrial wastewater discharges; establishing practices to control and remove contaminants from drinking water and to prevent its deterioration during storage and distribution; and assessing the nature and controllability of releases of toxic substances to the air, water, and land from manufacturing processes and subsequent product uses. This publication is one of the products of that research and provides a vital communication link between the researcher and the user community.

The seminar/workshop proceedings documented herein represent an effort to summarize current research and development activities in the United States, Canada, and Europe to improve aeration system design, operation, performance, and testing. This information will find direct application in the search for more cost effective wastewater treatment systems and more energy efficient operation. The brief workshop summaries presented at the end of these proceedings provide insight into the further research needs of the profession for this important unit process.

Francis T. Mayo, Director
Water Engineering Research Laboratory

# PREFACE

In the last few years, substantial effort has been devoted to the search for more energy efficient wastewater treatment systems. Since the aeration of wastewater represents one of the more energy intensive operations in wastewater treatment, research and development efforts have been aimed toward improvements in design, manufacture, testing, operation, and control of these systems. Among the research and development programs that are currently completed or underway are

- o clean water oxygen transfer test analyses and verification,
- o in-process oxygen transfer test development and analyses,
- o alpha factor test procedure development,
- o oxygen sensor evaluations,
- o fine bubble diffuser O & M evaluations,
- o dissolved oxygen control strategy analyses,
- o cost-benefit studies on aeration systems and aeration control, and
- o field-scale studies to optimize fine bubble design and operation.

The major goals behind these broad-based research and development activities are aeration system energy reduction and improved criteria for sizing oxygen transfer equipment. In extracting maximum performance from both traditional and emerging aeration approaches, four principal technical areas must be addressed: system design including specifications for compliance testing, shop and field testing, operation and related maintenance, and air supply control.

This Canadian-American seminar/workshop has brought together experts and practitioners in the field of oxygen transfer to review and discuss the current status of aeration system design, testing, operation, and control. Sufficient workshop time was provided to discuss in some depth new technology and conceptsin aeration system design and operation. Areas of agreement and disagreement were identified and research needs described.

The proceedings of this seminar/workshop will provide the profession with a current state-of-the-art document on aeration system design, testing, operation, and control. It is anticipated that future workshops will update this work and, perhaps, provide even more details in specific areas addressed in this meeting.

William C. Boyle, Project Coordinator
University of Wisconsin
Madison, Wisconsin

# ABSTRACT

The objective of this seminar/workshop for consulting engineers, planners, and other professionals was to exchange information on recent research and development activities in the United States, Canada, and Europe on the design, operation, control, and testing of aeration systems.

The 3-day seminar/workshop consisted of short, formal discussions followed by an afternoon workshop session. Papers and current technical data on selected topics were presented in several technical sessions including aeration system design, operation, and maintenance; aeration system control; specification and testing of aeration equipment; and translation of oxygen transfer data from clean water to process conditions.

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The program for this seminar/workshop was conceived and designed through the efforts of Nobert Schmidtke and Bruce E. Jank of Environment Canada, Richard C. Brenner of the U.S. Environmental Protection Agency, and William C. Boyle of the University of Wisconsin-Madison. The local arrangements and research agreement administration were provided by John T. Quigley of the Engineering Extension of the University of Wisconsin. The contributions of the seminar participants and authors during the 3 days of meetings in Madison, Wisconsin, are gratefully acknowledged.

## AN OVERVIEW OF OXYGEN TRANSFER SYSTEMS

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When I was first asked to present the opening remarks to this conference, I was both flattered and delighted. Flattered to be asked to address such an eminent group and delighted to have people sit still and listen to a dissertation on my favorite topic, oxygen transfer. The reality of the situation became obvious, however, as I sat down to actually prepare my remarks. In most cases, an audience knows a little about oxygen transfer, and even the most fundamental observations are accepted as brilliant insights into the process. Here, however, the absolute reverse is true. This audience is probably one of the best informed in the world in dealing with the topic of oxygen transfer. What then does one say to inform or inspire such a group.

The answer I found lies in the old expression concerning our ability to recognize and separate the forest from the trees. We, in this group, in our professional activities, deal daily with the so-called "trees" of oxygen transfer. In fact, many of us spend a good part of our time dissecting the leaves. Under such circumstances, we may from time to time lose sight of the bigger picture and of the role that oxygen transfer plays in the overall science of environmental management. I will therefore, endeavor in these remarks to place in prospective the role of oxygen transfer in our technology and in our society. I will also attempt to identify the intrinsic functions other than oxygen transfer that are performed by the aeration equipment and the importance of these functions to the treatment process.

### SIZE OF THE FOREST

A rather general survey of the data available on municipal and industrial treatment installations suggest that in the North American Continent, there are approximately 1.75 million installed horsepower of aeration equipment in place. The distribution between municipal and industrial use of equipment is approximately 55% municipal, 45% industrial. The equipment includes all varieties of aeration devices. Municipal systems tend to employ more diffused air systems where industrial applications tend to favor surface aeration systems. The approximate installed value of equipment presently in the field is between 0.6 and 0.8 billion dollars. The market sales per year, including both replacement and new systems, varies significantly, but appears to average about $80 million per year after peaking at above $100 million per year. Operating costs for power to run these various systems may be expected to be in the range of 0.6 billion dollars per year. If applications in Europe, Australia and Japan are included for consideration, the level of installed horsepower about doubles and the costs of operation increase to approximately one and one-half billion dollars a year. There are in the United States and Canada approximately 70 discrete manufacturers of aeration of equipment, each making the ultimate device for efficient transfer of oxygen.

The size of the worldwide market cannot be estimated with any accuracy at this point in time, but the data presented here is enough to set the prospective. We are dealing with a large and expanding market of significant economic size. Available data suggests that the overall transfer efficiency of oxygen transfer equipment in place is quite low. In fact, we probably have the potential to save about one-half of all the power costs presently employed for the transfer of oxygen if systems could be optionally designed, built and operated.

The fundamental requirement to begin on such a program of optimization is agreement on a method of measuring oxygen transfer so that one can obtain a valid comparison between systems. Development and acceptance of such a test protocol is one of the primary objectives of this workshop and hopefully will be one of its principal accomplishments.

## IMPORTANCE OF THE PROCESS

No process is more fundamental to the successful operation of an aerobic biological treatment system than is the transfer of dissolved oxygen. Unless dissolved oxygen is available where and when the bacterial system requires it, the process will not function. And, if adequate oxygen is available, the process will function almost in spite of all other upsetting conditions. Recent studies have shown that the absolute concentration of oxygen is also of significant concern and influences the type of organisms that will dominate and the general behavior of the process.

Oxygen transfer capability is fundamental to proper employment of the one hundred to two hundred billion dollars water pollution control program that is underway on this continent. Without adequate and correct oxygen transfer, our capital investment will fail to achieve the goals that have been set forth and, in fact, our society will be forced to live in an unacceptable environment. Oxygen transfer has a major impact on the way we live and on the quality of life we can expect.

## ASPECTS OF THE PROCESS

The total function of aeration equipment must be viewed when we talk about oxygen transfer. A perfectly efficient, properly designed transfer system must do more than simply dissolve gas in water. It must provide an environment which is optimum for bacteria to carry on our intended processes. As such, the device is called on for a much broader class of service.

The oxygen transfer device is the essential mixer which carries on all the basic mixing functions within the aeration tank. Mixing is required in several senses. A velocity must be introduced in the tank which is sufficient to hold the solids in suspension and provide sufficient shearing forces that the floc does not become overly dense. The mixing must also be gentle enough that the floc is not totally dispersed and the colonies created by the bacteria are maintained and allowed to function. Further, the mixing must transport the bacteria and the food into close proximation so that reaction can occur.

Another function of mixing involves the macro-system of the tank. The method of mixing will determine the degree to which the process approaches a plug flow or completely mixed system. While the nature of the system has distinct implication with respect to treatment, it also influences, in many cases, the energy required for oxygen transfer. Consider, as an example, the case where the raw waste exhibits a low $\alpha$, say 0.6. Whereas the treatment effluent $\alpha$ approaches 0.9.

If the system operates in a plug flow mode, the energy requirements for oxygen transfer will be approximately 25% more than that in a completely mixed system. A dispersion of the organics through the tank is principally affected by the mixing patterns induced by the aeration equipment.

A second point that must be considered by this conference is that oxygen transfer systems must achieve optimum transfer without a sacrifice of mixing characteristics. In fact, the design of all equipment must take into account mixing otherwise the function of the system will not be accomplished.

## FLEXIBILITY

Another major area that must be considered in designing optimum aeration systems is flexibility. The oxygen requirements of any biological system will vary widely during a typical day, week or month. Typical designs today often provide aeration equipment capable of meeting the maximum four-hour demand that occurs 10% of the time. If such a design is chosen, a system operating at that capacity will be capable of overaerating more than 95% of the time. Obviously, significant turn-down must exist both in the oxygen transfer and in the mixing capability of the system.

Many of the systems in place today that achieve high efficiency do so at a significant sacrifice of flexibility. Surface aeration systems are very difficult to design in such a manner that a significant number of units can be taken out of service. Equally, diffused aeration systems have an absolute limit on the minimum flow rate at which the necessary roll and hydraulic dispersion takes place within the tank. Clearly, this lack of flexibility has a significant impact on energy efficiency.

To install a device that is capable of transfer oxygen at the three or even four pounds of oxygen per horsepower hour and then to be forced to operate the device at double or triple its required capacity, a substantial percent of the time can reduce the overall oxygen transfer efficiency observed by the client to a very low level. The lack of flexibility, combined with the lack of sensing and control systems that allow operators to actually adjust oxygen levels within the tank, is probably the single largest contributor to energy waste in the field. A small amount of attention has historically been paid to this very important problem.

Emphasis then, must be placed not only on providing a device that under some ideal condition of flow, geometry and like can transfer oxygen in an efficient mode, but also to the development of a system that can maintain the efficiency over the wide range of conditions that would normally be encountered in an operating system.

This conference should address this requirement as one of the significant concerns in oxygen transfer. Consideration to examining a range of operating conditions should also be given in any test method so that the efficiency of an aeration system can be defined over the normal operating range of the system and not at the single point at which it is most efficient.

## FACTORS RELATED TO THE LIQUID

Investigators have delineated the influences of oxygen saturation, temperature and organic components on the ability to transfer oxygen. The coefficient β (beta) has been defined to estimate the approximate deterioration of saturation in the presence of various inorganic components while θ (theta) is used to estimate temperature impacts and the coefficient α (alpha) has been used to estimate the effect of organic compounds on the basic transfer mechanism itself.

To date, the work concerning oxygen saturation has been quite good and although there is no absolute agreement on the coefficient β and all the factors that effect it there are generally acceptable techniques and estimating procedures available for this coefficient. Similarly, although not all totally accept the mechanism for estimating the impact of temperature, the range of variance between the investigators is narrow and a reasonable estimate of this variance can be made. The major remaining questions concern themselves with alpha.

Probably in no area of science in which I am involved do we more behave like the blind men examining the elephant than in the measurement of α. Each set of experiments and each investigator seem to create conditions where a particular influence on oxygen transfer is observed. I have in my own experimentations seen alphas ranging from 0.3 to 1.9. I have seen changes in the physical conditions of my experimental system, change the measured results by 40% to 50%. I have seen people I consider competent investigators perform the same experiments as I did and get significantly different results. These variances and frustrations have been experienced by almost all who have worked extensively in this area. To this date, we do not have a reasonable theory for the impact of organic materials on oxygen transfer. What is necessary, is the development of some basic thesis which explains the behavior and allows us to compare our experience on a meaningful scientific basis.

It is likely that organic agents influence oxygen transfer in at least two ways. First, by collecting at the liquid-gas interfact and influencing diffusitivity and second, on influencing the very nature and size of the transfer surface. There may, in fact, be other more subtle effects including minor changes in viscosity, pumping ability and the like which further influence the overall transfer of a device.

It should be a goal of the group meeting here to attempt to develop a basic thesis relating to the impact of organic materials on oxygen transfer, so that the scientific community can compare their experiments to that thesis and provide criticism and input that will lead to the evolution of an acceptable understanding of the phenomena.

## SUMMARY

Oxygen transfer should be seen as a major industry in North America and in the world. Significant monies and energies are being expended to carry out this most important process. The protection of our environment is highly dependent upon the successful implementation of this process. We, gathered here, to a large degree, have responsibility for seeing that continual progress is made in more efficiently employing these resources and reliably carrying out the intended process. Several major problems face us - we must design flexible, efficient systems under a variety of ever-changing conditions. We must better define technology and the methods of testing that are to be applied. We have before us a significant challenge, but we have around us an elite group of skilled scientists who I am sure are up to the challenge.

## DISCLAIMER

The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency, and no official endorsement should be inferred.

# DIFFUSED AERATION - TYPES AND APPLICATION

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In the 1982 product guide section of the Journal, WPCF, (1) under the heading "Diffusers, Air, over 30 suppliers are listed. The list is probably not all inclusive and many of the suppliers listed offer several types, my company for example, offers four types. So, the variety of available air diffusion devices is likely many times 30.

It has been customary to suggest the relative clean water performance characteristics of the devices by categorizing them as fine bubble (best ?), coarse bubble (good ?) or even medium bubble (better ?). This categorization is somewhat subjective and relative and in some cases misleading in terms of field performance. MOP 5 (2), (3) and MOP 8 (4) distinguish only two categories; porous and non-porous. Porous diffusers are defined as being constructed of either natural ceramic or synthetic media and non-porous are described by specific example.

Porous diffusers are available in square or rectangular flat plates, round discs either flat or profiled, domes and tubes. Non-porous diffusers are available in a wide variety of shapes and materials. They are constructed of metal such as galvanized steel or stainless steel, a variety of plastics, rigid or flexible. They can be oriented horizontally or vertically, have single or multiple air release ports, fixed or variable area air flow passages, baffled or plain.

Several types of systems are a combination of air diffusion and mechanical mixing. Systems such as jets, submerged turbines, and mechanically rotated air curtains or baffle walls are examples.

The general function of all the devices in activated sludge aeration applications is to:

- Transfer oxygen
- Promote mixing of the oxygen with the aerated solids and suspend the solids
- Promote flocculation
- Resist deterioration of performance with time.

A brief review of the performance history of U.S. aeration practice gives some insight into the reasons for the multitude of devices.

Oxygen transfer theory and early clearwater testing indicated that porous type diffusers are relatively more efficient than other types. Thus, early systems consisted of flat, square or rectangular porous plates. Plates were typically constructed of fused or bonded aluminum oxide or silica. They were installed in fixed plenums on the tank floors, one plenum accommodating multiple individual plate units. Drop pipes supplied air to the plenums. Adjustment of air supply among the plenums was possible, but within a plenum air was distributed among the plates in an inverse relationship to their resistance to flow.

The operating history of such systems indicated a wide range of service life, but generally there was a loss of performance with age or time in service (2) (3).

Loss of performance was characterized by system pressure increase, a decrease in oxygen transfer efficiency, or both.

Loss of performance was attributed to fouling or clogging, air side or liquid side.

A variety of methods to attempt restoration have been tried, with varying degrees of success.

The general lack of sustained success of those cleaning methods and the expense of the techniques relative to the then prevailing cost of power, encouraged the development of more clog resistant systems. Similar development proceeded toward systems that could be removed from a tank without draining, to make cleaning easier and non-process interrupting.

Porous tubes or non-porous tube type diffusers were developed to go along with the removable header systems. The tubes consisted of a diffusion element and detachable mounting hardware for connecting to the air header.

Porous tubes were constructed of material similar to the porous plates or bonded glass, synthetic beads, etc. Non-porous tubes were available in many varieties, including woven socks, wrapped cords, perforated polymers, etc. Of necessity the aeration equipment was located near the side or sides of the tank, a position which clear water tests showed to be less favorable than other locations (5) (6). To varying degrees the tube type diffusers were generally vulnerable to the same operating problems as the porous plates.

Developments continued to be in the direction of an economical, non-clog, efficient, low maintenance system.

Much recent work (6) (7) (8) (9) (10) (11) has been done in an effort to define and refine the operating characteristics of such devices.

More or less, the variety of devices available today were developed to overcome the historical problems associated with high performance systems.

The recent change in the relationship between power costs and maintenance costs has caused a renewed interest in systems that obtain relatively high oxygen transfer but require relatively more maintenance than others.

The starting point for system performance comparisons has traditionally been clear water tests. Recent advances toward a standard procedure (12) for clear water testing and data evaluation will provide increased information on the relative performance characteristics of the various systems.

Clear water transfer data of the several systems vary quite widely but can be characterized in a few general ways. For a fixed condition of submergence, and diffuser placement:

- Some diffusion systems decrease in OTE with increased applied air or power/volume.
- The mass of oxygen transferred increases, but less than proportional to applied air rate.

- Some diffusers exhibit more or less constant OTE with applied air rate, and the mass of oxygen transferred is proportional to applied power.

- Some systems increase in efficiency with an increase in applied power. For these the mass of oxygen transferred is more than proportional to applied power.

- Some systems exhibit an optimum point of operation or a concave downward performance curve of OTE vs. air rate. From the optimum point, the mass of oxygen transferred also increases with increase air, but less than proportionally.

Field performance is presumed to be related to clear water by the application of appropriate conversion factors such as alpha, beta, theta.

In the past, attempts to determine the relationships between field and clear water have yielded unclear and inconsistent results. Recent activities (13) in improving field evaluation techniques show promise of being able to determine the relative performance of the various devices in service.

With the renewed interest in porous diffusers, new techniques have been developed to characterize and the performance and refine the design of the diffusion element (14). A discussion of the techniques is included in the agenda of this conference.

The characteristics can also be used to monitor the field performance of the system at the diffuser. Monitoring the diffusers in service aids in evaluating performance with time and gives an early indication of the necessity for taking corrective measures or cleaning, if required. Comparing the characteristics after cleaning to new, gives a direct indication of the effectiveness of the cleaning technique.

One of the most significant system differences between currently available porous grid systems and past systems is the incorporation of an individual flow control orifice for each diffusion element. Characteristically a plot of air flow vs. $\Delta$ P for the wetted media only, is very flat. A variation from average of ± 10% in specific permeability can result in a 200% range in air flow. The flow control orifice compensated for this and assures almost equal air flow among the diffusers in a grid.

Other developments relate to improving air flow uniformity or air flux rate across an individual element. Uniformity of air flow results in more efficient use of the diffuser's active area and has the potential for more uniform restoration of performance upon cleaning.

Another recent development is a method of in-place non-process interruption cleaning (16). The process was designed primarily to restore performance after plugging due to inorganic scaling. On pilot scale tests it has also demonstrated some effectiveness in restoring performance after other types fouling also (15).

Recent field operating data suggests the desirability of an inlet mix system, to reduce the effects of surface active agents and the suspected effect of high oxygen uptakes at the inlet. Other similar concepts suggest applying the inlet load to the tank over a long inlet weir (16).

Recent studies (7) show that the porous grid system can be designed to accommodate highly loaded processes. However, flow, or more appropriately load equalization is probably a more desirable feature for the porous grid system than for other types. An example of diurnal BOD load variation is shown in Figure 1 (17). The oxygen demand may be expected to vary in a similar sinusoidal manner. Dividing the curve into areas above and below daily average shows a greater than 4:1 variation in demand.

Since the transfer rate of porous grid is less than proportional to air rate, a 4:1 load variation requires a greater than 4:1 air application rate.

Also significant is that 74% of the total daily load is being transferred at a rate above the average daily rate.

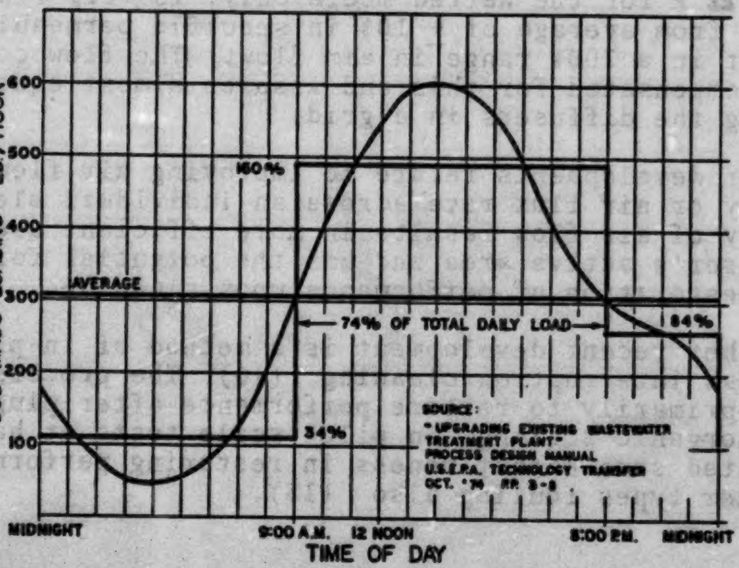

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FIG. I

EXAMPLE OF DIURNAL LOAD VARIATION TO A U.S. W.W.T.P.

Some other types of systems do not require such a wide range in air flow capability and would perhaps be more efficient than a porous grid under peak loading conditions.

Non-process considerations are also important to the successful long term operation of any system. These include characteristics of the liquid environment, such as temperature and ranges of change in temperature, corrosion or scaling tendencies, presence of chemical constituents which may degrade system components; reliability and cost of power; O & M skills and cost relative to power; and climate.

In addition to activated sludge aeration, other unit processes employ diffused air. Some examples are pre and post aeration, liquid transfer channels, sludge digestion, sludge conditioning, aerated grit chambers, flow and load equalization tanks, and transfer of other gases such as ozone.

Presently a great interest exists in high performance systems, but there appears to be sufficient variety of applications that no single system is likely to evolve as the universal system of choice.

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## DISCLAIMER

The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency, and no official endorsement should be inferred.

# MECHANICAL AERATION SYSTEMS - TYPES AND CHARACTERISTICS

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## ABSTRACT

Many types of mechanical devices have been invented for the purpose of aerating liquids. These devices are categorized and their performance characteristics summarized.

While it is possible to rank various aeration devices in order of energy efficiency under standard test conditions, the optimum selection of aeration equipment for a specific application must take into account other factors. Waste water characteristics, oxygen up take rate and turndown ratio, basin depth and climatic conditions can each have a significant impact on the final selection of the aeration system.

## INTRODUCTION

In the waste water treatment process, the consumption of soluble organics by aerobic microbes requires the dissolution of large quantities of oxygen into the mixed liquor. Because the solubility of oxygen in water at atmospheric pressure is limited, the driving force is small. These circumstances dictate the need for a large interfacial area in order to achieve a mass transfer rate which will satisfy the process demand.

The only other means of increasing transfer rate is to increase the driving force. This can be accomplished by conducting the mass transfer operation at higher pressure or by using an oxygen enriched gas supply. In either case, any increases in the gas-liquid interfacial area which can be accomplished by the aeration device will provide proportionally higher rates of mass transfer. Therefore one essential goal of any aeration system must be to maximize the gas-liquid interfacial area.

Another consideration in the design of an aeration system is the mixing capability of the aeration device. The direct pumping capacity plus the induced flow must be capable of maintaining a uniform suspension of biological solids and a positive dissolved oxygen concentration throughout the aeration basin. Mixing can lengthen the duration of contact time between the bubble and the liquid and thus increase the oxygen stripping efficiency. Rapid blending of toxic or other feed upsets will dilute their impact on the biological process.

Moderate fluid shear rates can minimize the negative affect of surfactants on mass transfer rates. The velocity of the fluid relative to the bubble produces a washing effect at the interfacial film. Evidence of this phonomenon has emerged in "dirty water" test results which indicate different alpha factors for different aeration devices operating in the same waste water.

Not only do different aeration devices exhibit differing transfer efficiencies, circulation rates and fluid shear rates, but different biological process designs place different requirements on the aeration system. A high rate system treating high strength industrial waste in a relatively small aeration basin obviously places a different requirement on the aeration system than does a low rate aerobic lagoon system treating a dilute municipal waste. Similarly, different aeration devices have a different impact on the thermal balance of the process and the performance of the process may be either enhanced or hindered by the choice of aeration device.

## TYPES OF AERATION DEVICES

All aeration devices can be categorized as follows:

(1) Diffused air.

- (a) Coarse Bubble Diffusers.
- (b) Fine Bubble Diffusers.
- (c) Tubes or "Guns".

(2) Mechanical Aerators.

- (a) Low speed surface aerators (LSSA).
- (b) Motor speed surface aerators (MSSA).
- (c) Brush type surface aerators (BA).
- (d) Down draft aerators (DDA).
- (e) Submerged turbine aerators (STA).
- (f) Submersible aerators (SMA).
- (g) Jet aerators (JA).

(3) Special Aeration Devices.

- (a) Trickling filters.
- (b) Rotating biological contactors.
- (c) Miscellaneous proprietary devices.

This paper will be concerned only with Mechanical Aerators.

## CHARACTERISTICS OF MECHANICAL AERATORS

### Low Speed Surface Aerators (LSSA)

A surface aerator is one which has an impeller operating at, or near the free liquid surface. It produces an umbrella shaped discharge of liquid into the atmosphere, creating a large interfacial area through which the mass transfer of oxygen takes place. In the gas-liquid mixture, the gas is the continuous phase and the liquid the discontinuous phase. The supply of oxygen is infinite. The transfer rate is limited only by the rate at which the impeller can expose new liquid interfaces to the atmosphere.

"Low Speed" means that a speed reduction takes place between the motor and the impeller, usually by means of a gear type reducer. Operating speeds are typically in the range of 100 RPM in the smallest sizes and down to 35 RPM in the largest sizes. Low speed aerators employ large impellers, typically 4 ft diameter at 5 HP and up to 12 ft diameter at 150 HP.

LSSA's produce extremely high pumping capacities. For a given expenditure of energy, a lower operating speed requires a larger diameter impeller and the net result is an increase in pumping capacity. A speed decrease of 25% necessitates an increase of 18% in the impeller size which in turn produces an increase of 25% in the direct discharge capacity of the aerator. A 5 HP LSSA will typically have a direct discharge capacity of 13,000 gpm while a 150 HP unit will pump over 150,000 gpm. Total flow developed in the basin is typically three to six times the direct discharge rate due to entrainment and induced flow.

Due to their size and weight, LSSA's are usually fixed mounted but can be float mounted where necessary. (Refer to Figure 1.)

Advantages

- LSSA's are simple in design, easy to install and to maintain as they have no submerged components.

- Their transfer efficiency is relatively high.

- Their operation and maintenance costs are relatively low.

- Their high pumping capacity allows them to provide adequate mixing throughout very large volumes.

- A turndown of 30% to 50% is attainable with liquid level sensitive impellers and long term increases in plant loading can be accommodated by the use of adjustable blade tips.

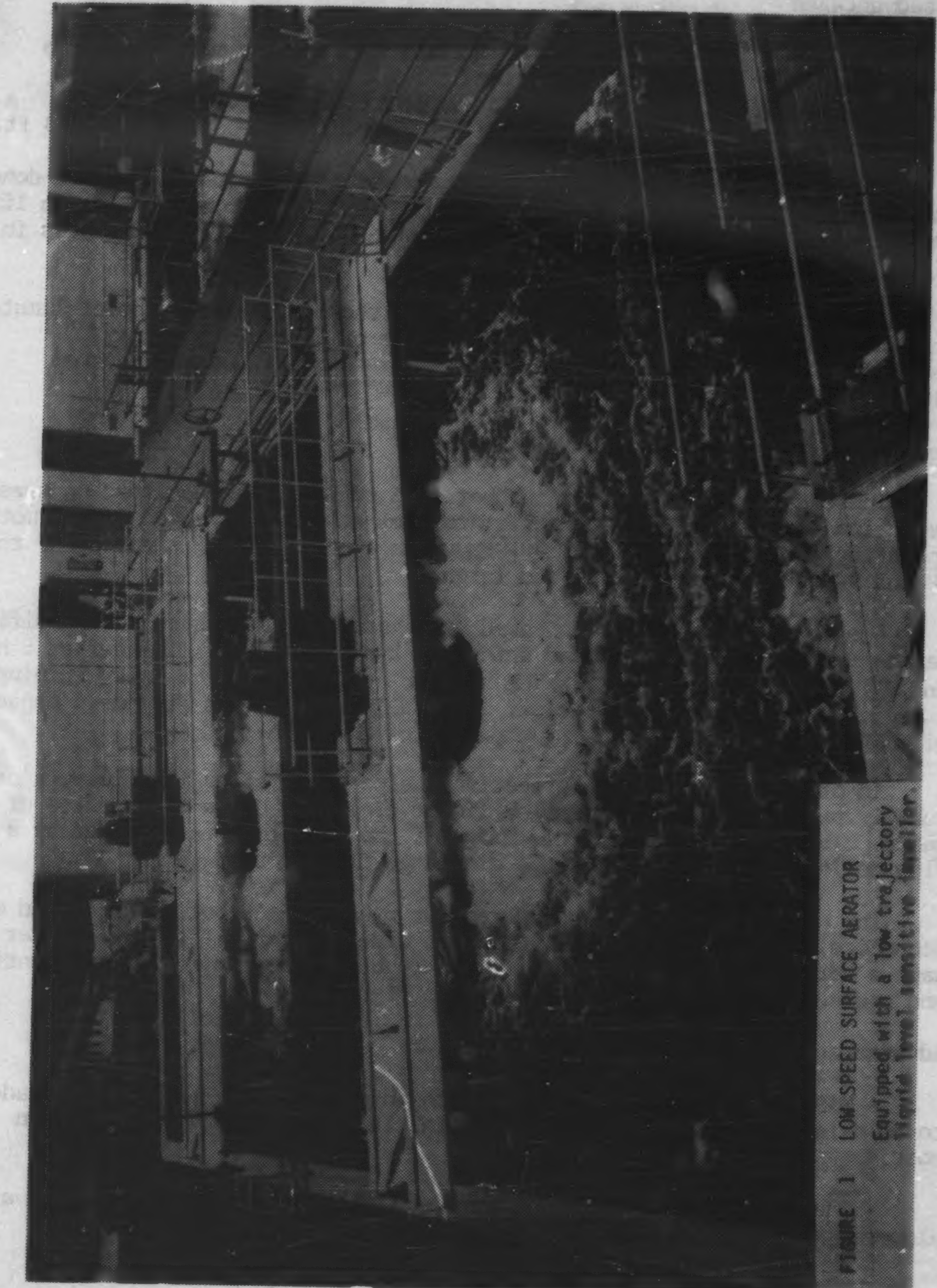

FIGURE 1 LOW SPEED SURFACE AERATOR

Equipped with a low trajectory liquid level sensitive impeller

Disadvantages

- LSSA's using only a surface impeller can provide mixing only to limited depths. The depth limitations range from about 10 ft for the smaller sizes to 17 or 18 ft for the largest sizes. LSSA's fitted with a lower mixing impeller or a draft tube are suitable for depths up to 30 ft.

- The turndown in transfer rate and power consumption can not be done independently of pumping capacity and mixing. Oxygen uptake rates can limit the system design when dealing with exceptionally high strength wastes in a high rate system.

- LSSA's can be designed to operate in the most severely cold climates when mounted on a properly designed platform or fitted with spray deflectors. The disadvantage is that heat loss can reduce the system kinetics.

## Motor Speed Surface Aerators (MSSA)

These devices are similar in operation to LSSA's except that no speed reduction is provided. The impeller rotates at the same speed as the motor. Operating speeds are usually 1800 RPM up to 10 HP, 1200 RPM from 10 HP to 40 HP and 900 RPM in sizes 50 HP and larger.

Due to the high operating speed, the impeller size is small. Modified marine type propellers are used which are typically 8 in. diameter at 5 HP and 24 in. diameter at 75 HP. The smaller impellers have a lower pumping capacity for any given motor size. Typically, the direct discharge capacity of a MSSA is 25% of the pumping capacity of the LSSA.

In order to improve the effectiveness of the smaller impellers, a draft tube is provided which draws mixed liquor from well below the surface of the basin. The vertical upwards discharge of the impellers is diverted by a flow deflector which controls the discharge trajectory.

The elimination of the speed reducer makes MSSA's both lighter and more economical than LSSA's for a given motor size. Because they are lighter and have a very limited shaft length, they are better suited for float mounting and are seldom fixed mounted. (Refer to Figure 2.)

Advantages

- MSSA's are portable and simple to install. They can be easily added to an existing system. They have low initial cost and low installation cost.

- There are no submerged fittings and no auxiliary equipment such as blowers.

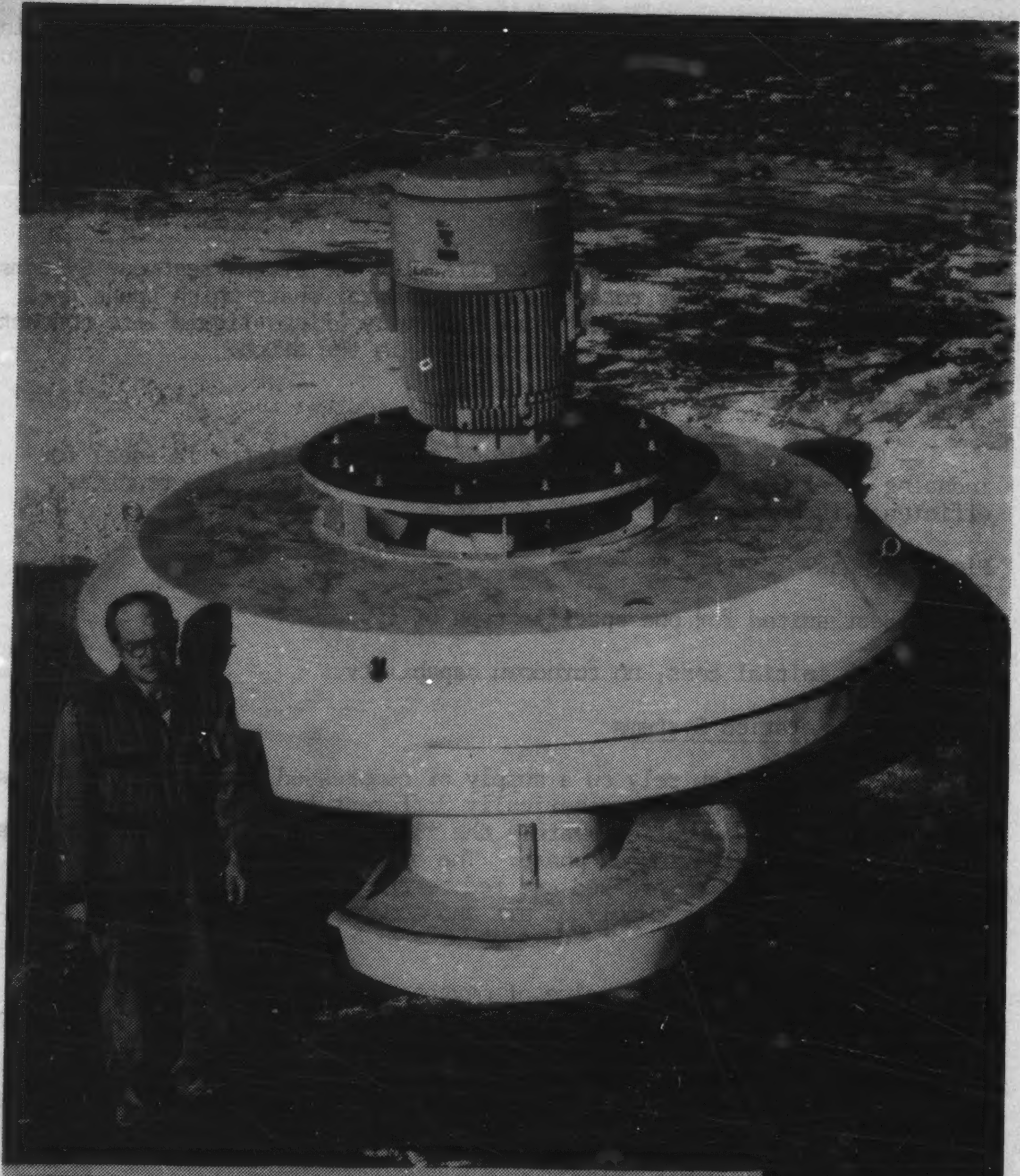

FIGURE 2 MOTOR SPEED SURFACE AERATOR

Equipped with ballast stabilized float assembly.

Disadvantages

- MSSA's have a relatively low transfer efficiency and a lesser mixing capability compared to LSSA's. There is no turndown capability other than turning off individual units.

- Cooling of the waste may decrease system kinetics. Flotation stability requires special consideration when installed in cold climates.

## Brush Aerators (BA)

These devices were developed specifically for Ditch Treatment Systems. The "Brush" type impeller rotates on a horizontal shaft which spans the width of the channel. The brush simultaneously brings liquid into contact with the atmosphere and induces circulation in the ditch.

Advantages

- The uni-directional discharge of the brush aerator is ideal for inducing circulation in a ditch or channel type system. The transfer efficiency is comparable to that of LSSA's.

Disadvantages

- Best suited for one specific type of system.

- High initial cost, no turndown capability.

## Sub-Surface Aeration Systems

All systems which rely on a supply of compressed air introduced below the free liquid surface are generically different from surface aerators. In these "sub-surface" aeration systems the fluid becomes the continuous phase and the gas the discontinuous phase. The oxygen supply is limited by the rate at which air is compressed and fed to the system. The transfer rate is dependent upon both the supply of air and the oxygen stripping efficiency of the aeration device.

Because the saturation value of oxygen in water increases with pressure, the mass transfer driving force increases with liquid depth. The stripping efficiency is observed to be significantly greater at greater basin depths thus reducing the air flow requirements. However, the increased discharge pressure required from the compressor tends to reduce the energy savings associated with reduced air flow requirements.

## Down Draft Aerators (DDA)

These units consist of a vertical shaft driving an impeller which is located inside a draft tube in the upper regions of the basin. Compressed air at very low pressure is injected into the draft tube immediately below the impeller. The down pumping impeller provides circulation throughout the basin and the downward velocities inside the draft tube carry the gas bubbles to the bottom of the basin. The fine bubbles provide a large interfacial area and the depth of the basin results in an elevated partial pressure of the oxygen. The net result is a stripping efficiency in the order of 40-50% of the available oxygen in the compressed air.

The combination of the lower air flow requirement and the lower static pressure at the point of injection results in a considerable reduction in the size of the blower. Consequently they are more energy efficient than submerged turbine aerators. Because they do not require an extremely long shaft they can operate at higher speeds than the STA which in turn reduces the torque and results in a smaller, less costly drive assembly. (Refer to Figure 3.)

Advantages

- High stripping efficiency, small blower low discharge pressure, high efficiency.

- High pumping capacity, ideal flow pattern for suspension of solids.

- Oxygen transfer and energy consumption can be turned down independent of mixing.

- No depth limitation.

- No icing, spray or misting problems in cold climates.

Disadvantages

- Draft tube increases installed cost.

- Less efficient in shallow basins.

## Submerged Turbine Aerators (STA)

A gas dispersing impeller is located in the lower regions of the basin and is driven by a vertical shaft. Compressed air is fed below the impeller and is dispersed by the hydraulic shear created by the impeller discharge. The resulting fine bubbles are transported throughout the basin and provide a large interfacial area and a long bubble detention time to facilitate the mass transfer rate required.

The fluid forces acting on the long overhung shaft and the critical speed considerations both dictate a low operating speed, thus requiring a speed reducer. The rotational speed of the impeller is generally in the range of 50 to 100 RPM.

The impeller diameters and the resulting pumping capacities are slightly less than those of LSSA's. However, they do afford the ability to independently control oxygen input without affecting the mixing capability

FIGURE 3 DOWN DRAFT AERATOR

of the device. This can be accomplished by operating the aerator at constant speed while increasing or decreasing the supply of compressed air. (Refer to Figure 4.)

Advantages

- No cooling of the waste or icing.

- Energy and oxygen transfer can be turned down independently of the mixing.

- Suitable for deep basins.

- Nearly unlimited oxygen uptake rate capability.

Disadvantages

- Requires blowers and submerged piping.

- Deep basins involve higher civil costs.

- Lower transfer efficiency.

## Submersible Aerators (SMA)

A submersible aerator employs a submersible motor to drive an impeller located near the bottom of the basin.

The impeller may operate in a draft tube or in a pump casing. It may operate at motor speed or the speed may be reduced through a submersible gear box. In either case, the impeller generates flow and compressed air is piped to the unit to be dispersed and circulated through out the basin. (Refer to Figure 5.)

Advantages

- High pumping capacity and good mixing due to use of speed reduction and larger diameter impellers.

- Independent turndown of energy and oxygen transfer without reducing mixing.

- Good transfer efficiency.

- No problems with cooling, spray or ice.

Disadvantages

- Requires a blower. Must maintain submersibe motor.

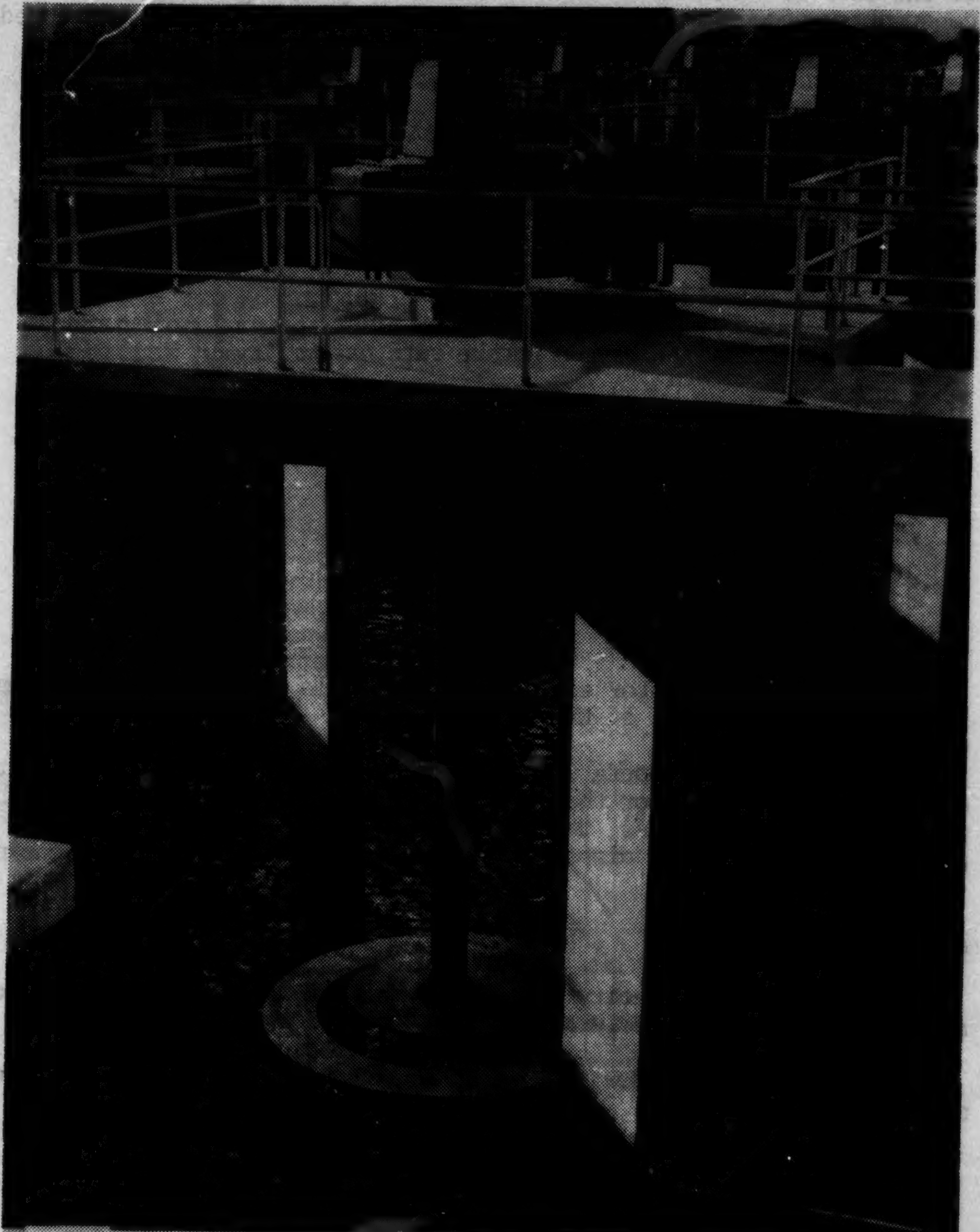

FIGURE 4 SUBMERGED TURBINE AERATOR
One of 72 similar units at Green Bay, Wisconsin.

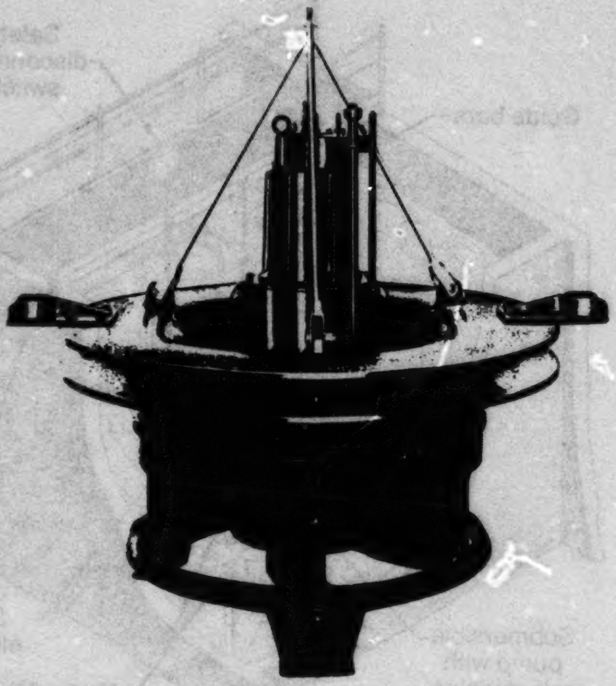

FIGURE 5 SUBMERSIBLE AERATOR

## Jet Aerators (JA)

Instead of a rotating impeller to aid in the dispersion of the compressed gas, these units utilize a venturi tube through which mixed liquor is pumped. The gas is admitted into the throat of the venturi at the point of highest velocity and lowest hydraulic pressure. This promotes a rapid dispersion of the incoming gas and the hydraulic flow pumped through the venturi provides the circulation necessary to achieve satisfactory mixing.

The air supply is normally from a centrally located blower room. The hydraulic flow through the venturi can be generated by either a remote dry well pump or by locally located submersible pumps. (Refer to Figure 6.)

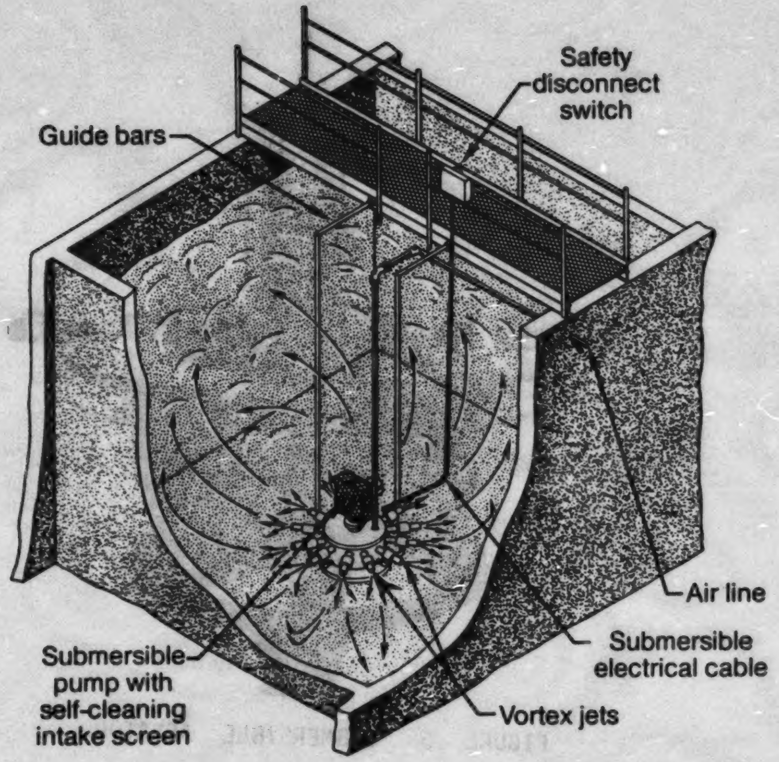
Safety disconnect switch
Guide bars
Air line
Submersible electrical cable
Submersible pump with self-cleaning intake screen
Vortex jets

FIGURE 6 JET AERATOR

One of several configurations available with or without submersible motors.

Advantages

- No cooling or ice problems in cold climate operation.
- Can turndown energy and oxygen transfer independent from mixing.
- Can be adapted to all types of basin geometry.

Disadvantage

- Requires blower and submerged piping.
- Relatively high installation cost.
- Must maintain submersible pump.
- Transfer efficiency comparable to STA.

## MIXING

Mixing continues to be a poorly understood yet very important aspect of aeration system design. Some of the misunderstanding stems from textbooks on the subject of biological treatment written by authors who had little knowledge of the technology of fluid mixing. In the absence of any better criteria, these textbooks suggest guidelines based on one parameter only, energy input per unit of volume.

Mixing requires that flow streams be developed within the basin. The greater the flow, the faster the mixing. Flow development can be either forced and/or induced. Some aeration devices develop high direct discharge rates while some rely entirely on induced flow. Even within the generic category of surface aerators there exists a 4 or 5 to 1 range in the discharge rates per unit of energy. Such wide variations in performance make any guideline calling for a minimum value of HP/1000 $ft^3$ too general to be of much value.

Flow generation capacity per unit of basin volume or "Turn-over rate" is much more pertinent, but still does not define the quality of mixing. A small diameter high velocity stream and a large diameter low velocity stream may have the same mass flow rate but they will produce different mixing results. Similarly, having all the mixing energy injected at one point in the basin will produce a different result than having the energy injected at multiple points. The direction, location and number of streams developed will have a significant affect on the resulting fluid regime in the basin.

Due to the difficulties in specifying equipment size or energy levels to assure adequate mixing, it makes more sense to specify the mixing performance required. Velocity specifications are popular but velocity is less definitive of mixing performance than one may be led to believe. For example, if the entire basin were rotating in the same direction at a specified velocity, there would be little mixing. The development of random mixing requires varying velocities and eddy currents so it is quite possible to find a zero velocity reading at some point in a well mixed basin. Nor is the measurement of dissolved oxygen a technically correct measure of mixing. Oxygen is a consumable commodity and the concentration must diminish the further the flow travels from point of injection.

Consideration of these facts suggest that the most practical and effective measure of mixing in an aeration basin is the degree of uniformity of solids distribution throughout the basin. Solids suspension is a basic process requirement and the fluid motion necessary to achieve a reasonably uniform suspension will be more than adequate to distribute dissolved oxygen. A reasonable mixing specification might call for a uniform suspension of biological solids throughout the basin within ± 10% of the average. Samples should be taken at several different depths in each of several locations in the basin.

## CONCLUSION

There are many generic types of aeration devices available, each having its own peculiar characteristics. Important variables include mixing, alpha factor, basin geometry constraints, ease of installation and maintenance, suitability for cold weather operation, initial cost, energy efficiency and turndown capabilities.

Various process designs place different demands on the aeration system in terms of mixing and oxygen uptake rate. Site conditions may dictate the use of deep or shallow basins. Climatic or aesthetic concerns may dictate the use of covered basins. High peaking factors may require a large turndown capability.

It is incumbent upon the treatment plant designer to select an aeration system best suited to the needs of that particular project. An awareness of the characteristics of all generic types of aeration devices is necessary in order to arrive at an optimum selection.

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# OXIDATION DITCH AERATION SYSTEMS - TYPES AND CHARACTERISTICS

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## ABSTRACT

During the last ten years, the oxidation ditch aeration system has gained in popularity. Various types of basin configurations have been proposed and used, along with numerous types of aerator/mixing devices; the advantages and benefits claimed with these widely varying configurations and aerator/mixing devices have been very widespread and sometimes conflicting in nature, especially concerning advantages of oxygen transfer efficiency and nitrification/denitrification capabilities. The low-loaded activated sludge treatment modes associated with these systems do achieve consistent degrees of high treatment performance. Oxidation ditch systems do offer, because of their continuous flow circuits and uni-direction flow pattern, low-energy mixing requirements, resulting in more turn-down flexibility for both the small and large size treatment plants.

## INTRODUCTION

The oxidation ditch aeration system, a treatment mode chosen more and more frequently in recent years for a wide range of treatment applications, is typically classified as a low-loaded, looped-reactor, activated sludge system where aerators of a specific type provide oxygen and establish an uni-directional mixing pattern to the basin contents. While designs have been proposed for plants over 100 MGD in the United States (and are in operation for plants over 100 MGD in Europe) the system, because of its ideal mixing regime, is especially suitable and attractive for the small to medium size extended aeration plant where nitrification is required. Part of the systems current popularity is undoubtedly due to the favorable EPA publication released several years ago containing a survey of the operations and performance of a large number of oxidation ditches (1). According to this report, the oxidation ditch, when compared to alternate treatment system modes, showed a higher and more consistent performance level. These results, however, may be more due to the forgiving nature of low-loaded systems rather than the unique basin or aerator characteristics associated with looped reactor systems. Low-loaded conventional aeration systems, though lacking the mixing (and in many cases, flexibility) advantages of a looped reactor, should have the same degree and dependability of treatment performance, at least under normal influent conditions.

Developed in Holland in the fifties, the oxidation ditch was brought over to the United States in the sixties and successfully applied in various regions of the country. These early ditches were all single-loop basins, usually in a race track configuration with sloped walls and an elongated straightaway section, and employed horizontal rotary aerators, either of the cage or brush rotor type. Most of the brushes were manufactured by Lakeside or Passavant and used the trade names of Magna Rotor or Mammoth Rotor. The orbal system, a multi-channel oxidation system using aerator discs was introduced in the late sixties as were jet aeration channels (a single loop system) using subsurface jet aerators. Later the Carrousel System, a folded-loop reactor using slow-speed, large diameter, vertical surface aerators at the channel ends, were introduced. In all, there are at least seven variations of the oxidation ditch system in operation in the United States, with the beforementioned four being by far the most dominate. The list of ditch-type systems is as follows:

a) Single-loop systems using cage or brush aerators.
b) Orbal (multi-channel system using disc aerators).
c) Jet aeration channels.
d) Carrousel (folded-loop system using vertical surface aerators).
e) U-tube ditches.
f) Counter-current ditches using diffused aeration.
g) Burns & McDonnell Treatment System.

Each system will now be described in more detail.

## Single Loop System

There are presently more than three hundred single-loop oxidation ditch plants using horizontal rotary aerators in operation in the United States and Canada. While various types of horizontal rotary aerators have been tried, the dominate aerator used is the brush rotor, usually operating in the speed range of 50 to 70 rpm with an overall aerator diameter of approximately 42 in. The horizontal rotor of the brush consists of a 14 in. diameter support torque tube on which 14 in. long steel blades are secured. Their horizontal lengths can be as long as 30 ft.

While most of the early oxidation ditches were designed with shallow depths and sloped side walls, deep tank ditches with vertical walls have now been used. Many of the early ditches were for plants smaller than 1 MGD; low-loaded oxidation ditches are now being proposed for plants larger than 20 MGD.

High-loaded oxidation ditches have been used in the past for larger size installations. Their process results are very similar to conventional aeration systems. Since mixing is usually never critical for high-loaded systems, it would seem that oxidation ditches would not necessarily have strong advantages for this type of application. Care also must be taken in these types of designs that velocities established by the ditch aerators do not become excessive, resulting in a loss of aerator oxygen transfer capability.

Brush aerators have good oxygen transfer efficiency rates, being approximately 3 to 3½ pounds of oxygen/hp-hr at standard conditions for a wide range of rotor speeds and blade immersions. Mixing can also give a requirement of 21,000 gallons of basin volume for every foot of rotor in order to maintain proper channel velocities (2). Since actual channel velocity and mixing efficiency is dependent on the frictional resistance of the walls and floor, the shape and size of the ditch, and other mixing factors such as channel bends and constrictions, the 21,000 gallon figure is probably conservative for many of the larger ditch applications. Baffle deflectors, used immediately downstream of the brush aerator in the deeper ditches, offer a considerable flow constriction and significantly reduces the velocity; usually this is beneficial since the higher velocity rates can reduce the oxygen transfer capability of horizontal rotary aerators.

A loop reactor is a continuous flow circuit and this in itself lends to the mixing efficiency. A uni-directional mixing pattern obviously requires less energy than a haphazard mixing pattern in maintaining minimum velocity levels for solids suspension. In ditch applications where several brush or horizontal rotary aerators are located in the same basin, it is possible to remove one or more from service without creating localized mixing inadequacies such as would be found in conventional aeration tanks with mechanical surface aerators.

The blade immersion of the brush can vary from 5-14 in. to alter the oxygen delivery rate; rotor speeds of brushes can also be varied to alter the oxygen output. In many cases, this variance is only possible because of the mixing superiority offered by the brush aerator in an uni-directional flow basin; in other words, oxygen delivery can be lowered to match oxygen need without being penalized for mixing. Again, this is a strong advantage for low-loaded systems.

## Orbal

Over one hundred orbal oxidation ditch type plants using aeration discs are in operation in the United States. Developed in South Africa, the orbal is a multi-channel system where the degree of oxygen input can vary from channel to channel by means of the number of aeration discs provided in each channel. Considerable process flexibility is allowed and numerous medium-load activated sludge process modes (such as contact stabilization or step aeration) have been used. The extended aeration mode however, is the process mode most frequently used. This mode has the process advantage of nitrification/denitrification occurring within the same sludge system.

The aeration discs used are only ½ in. thick and 4 1/2 ft.in diameter. They are split in two semi-circular half sections for ease of removal or relocation on the shafting. Typical rotor speeds are between 45 to 60 rpm. Like the brush, the oxygen transfer efficiencies of the disc are good, being in the range of 3.0 to 3.4 pounds of oxygen/hp-hr at standard conditions. The aeration discs consists of ½ in. diameter aeration holes and numerous triangular protrusions which provide the oxygen and liquid propulsion. Despite the narrow thickness of the disc, it has a very efficient mixing capability; design figures used are 1 hp worth of disc aerators for 50,000 gallons of basin volume (3).

While each channel of the orbal system duplicates the characteristics of a loop reactor system, the kinetics of several complete mix channels in series has some pronounced differences over a single channel arrangement. The oxygen uptake rate, for instance, is most likely to vary considerably from channel to channel, with the influent channel having the highest rate and the last channel having the lowest rate. In this way, the orbal system approaches the characteristics of a plug flow reactor; the desired effect is to eliminate any premature exiting of incoming organic matter, especially the ammonia.

It should be mentioned that the single channel reactor is not a true complete mix system despite the rather quick dispersion of the influent flow over the entire contents of the tank. Actually the channel system has a "moving-belt" type circulation. The degree of complete mixing can be very dependent upon the design of the basin, especially the size. For a small channel where the detention time is long, an individual flow particle may, on the average, make over 500 circuits within the basin. A larger, more elongated channel, having the same 1 fps channel velocity, and a reduced detention time, will have considerably less circuits; as the number of circuits approach one, the characteristics of the system will approach that of a plug flow reactor.

With the orbal system, the basin configuration is usually kept oval on circular to assure a more complete mix characteristic of each channel. Usually the first or influent channel is largest in volume in order to gain the buffering advantage of complete mix.

The isolated zones of the orbal are ideal for achieving a consistent high-rate nitrification/denitrification performance

(4). In the orbal system, a high degree of process work will occur in the first channel resulting in a depletion of dissolved oxygen despite high oxygen input. Nitrification has been shown to occur in these oxygen depleted zones; because of the readily available carbon source, all of the nitrates formed in the first channel will be denitrified allowing the oxygen input of the discs to be reduced. The nitrification rate can be more than 90% complete in this oxygen-depleted or anoxic zone; complete nitrification for diurnal ammonia peaks is provided by the remaining channels where the dissolved oxygen levels, because of the low uptake rates, can be kept excessively high.

A high dissolved oxygen level can be kept in the last channel of an orbal basin without a significant energy penalty. Because of the efficient mixing provided by discs in a channel reactor, a low oxygen input can be provided; because the uptake rate is exceedingly low, the resultant dissolved oxygen level will still be quite high.

The zero dissolved oxygen level associated with the first channel is an important energy consideration because of the high driving force provided with oxygen delivery. In the first channel of the orbal, the dissolved oxygen level of the liquid passing into the disc aerators is zero while the liquid passing from the aerators also is near zero. This is accomplished by having wide-spread oxygen delivery application points of low intensity; the low intensity delivery rate coupled with the relatively high oxygen uptake rate of the liquid allows most of the delivered oxygen to be immediately used instead of being carried long distances in the channel flow in the form of increased dissolved oxygen levels. Keeping the oxygen delivery rate below the uptake rate allows this to take place in this channel.

Like the brushes, the discs operate with a variable immersion level, usually in the range of 9 to 21 in. Typically the immersion level is allowed to vary over this range over a 24 hour period; a submerged orifice and dual effluent weirs are used to achieve this purpose.

The maintenance requirements for disc aerators and brush aerators are similar. Frequent greasing of bearings supporting the horizontal shaft sections is a necessity. Disc aerator assemblies used with the orbal system usually have more bearings than the brush aerator assemblies used with the single channel ditch. The steel aeration blades of the brush are more subject to corrosion problems than are the plastic aeration discs.

Aeration disc equipment is manufactured by Envirex Inc. The orbal system is not a proprietary system and no license fees are charged for the use of the process. Figure 1 is of an orbal basin showing the varying design dissolved oxygen levels.

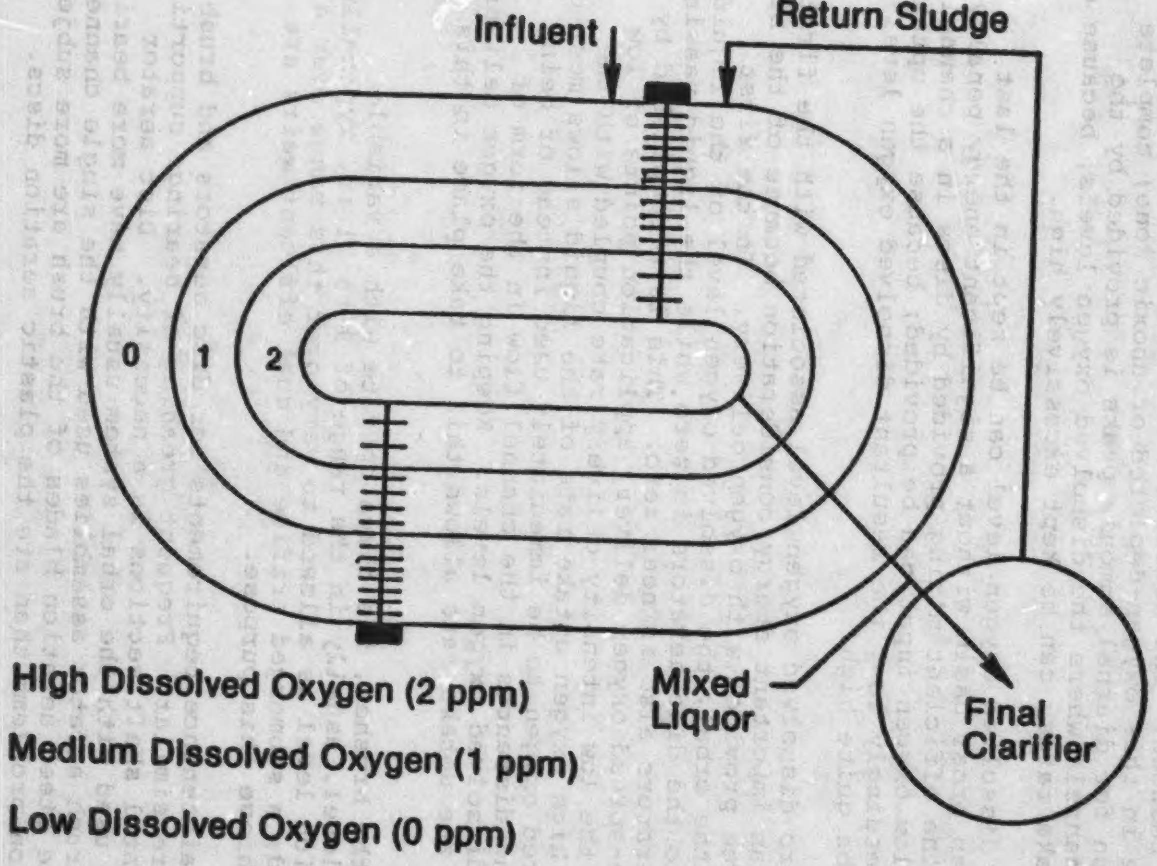
Influent
Return Sludge
0
1
2
Mixed Liquor
Final Clarifier
High Dissolved Oxygen (2 ppm)
Medium Dissolved Oxygen (1 ppm)
Low Dissolved Oxygen (0 ppm)

Figure 1. The Orbal System

## Jet Aeration Channels

Horizontal jet aerators have in the last decade been used successfully in a looped reactor configuration. These aerators use both pumps and compressors to propel and aerate the liquid contents of the ditch. Because the aerator is a sub-surface aerator rather than a surface aerator, the channel depths can be much deeper. This allows an advantage over the horizontal rotary aerator described in the previous two sections which for practical purposes, can only work, with depths between 5 ft. to 15 ft. Optimum depths for a jet aeration channel would be between 15 ft. to 20 ft. Deeper channels are an advantage in cold weather climates, allowing less heat loss.

The typical jet aeration device consists of a directional mix header with aerator nozzles on one side of the header. The header has a manifold containing separate air and liquid conduits which distributes each flow stream uniformly to the nozzles. Figure 2 shows the typical jet aerator device.

The jet channels have found numerous applications for industrial wastes where high uptake rates are a frequent design occurrence. The oxygen transfer efficiency is similar to that of diffused aeration systems with rates between 4 to 5 pounds oxygen/hp-hr being achieved in clean water. Since the device is a sub-surface aerator, it is probably subject to lower transfer rates in dirty water, making the efficiency rate more equal to that of a brush or a disc. However, the looped reactor regime in which it operates in, does give it a strong mixed efficiency advantage over diffused aeration devices in conventional tanks.

## Carrousel

The Carrousel System, developed in the Netherlands in 1971, has made a strong entry in the oxidation ditch market during the last few years in this country. Presently there are several plants over 5 MGD operating successfully.

The Carrousel basin uses a vertical shaft aerator with a large diameter impeller and a low operating speed, usually less than 30 rpm. The aerator is located at every 180° turn of the channel. Using units as large as 100 hp, this aerator unit would seem to have a cost advantage over the brush or disc aerator in that it involves a lot less machinery and does not span the full channel. Another advantage is that, while the Carrousel reactor might be quite new to the American consultant, the low-speed vertical aerator is of a generic type that has been familiar to the consultants in this country for a very long time.

Despite the numerous possible folds with the Carrousel basin (which evidently depends upon the size of the basin and the

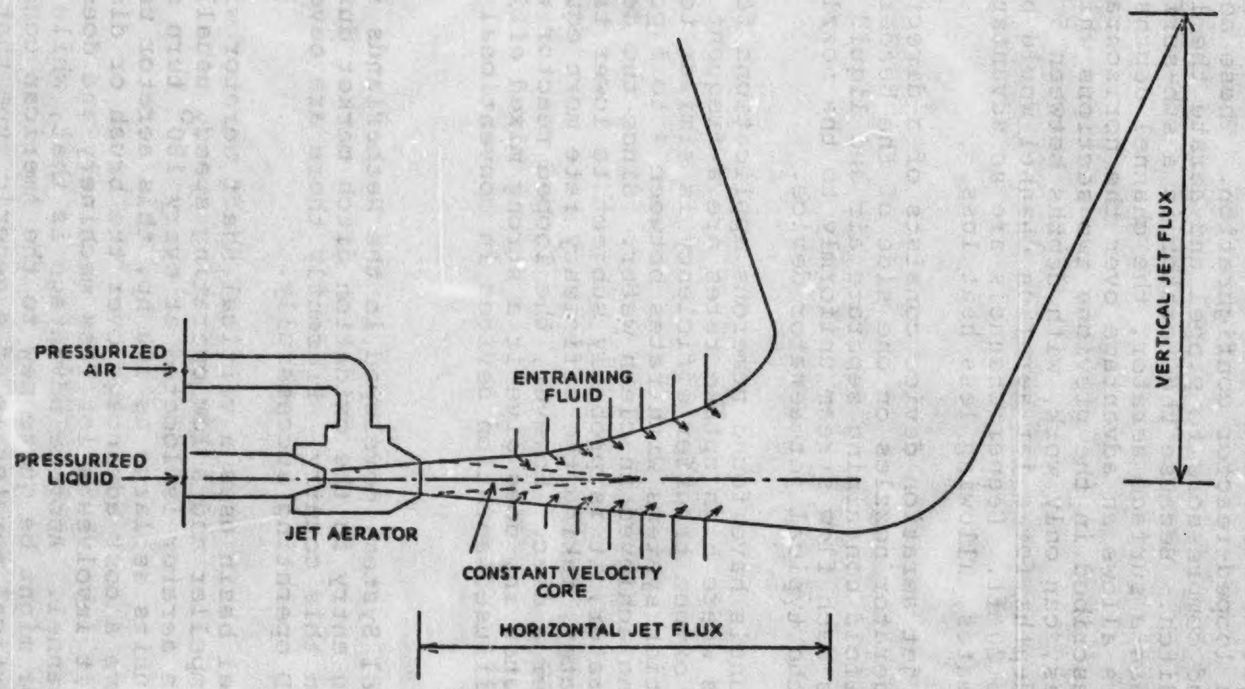
PRESSURIZED AIR
PRESSURIZED LIQUID
JET AERATOR
ENTRAINING FLUID
CONSTANT VELOCITY CORE
HORIZONTAL JET FLUX
VERTICAL JET FLUX

Figure 2. Jet Aeration Mixing Pattern

number of individual aerators to be used) the reactor is still a single loop system. An identical arrangement could be achieved with a brush oxidation ditch by elongating this ditch's straightaway section and bending it over. An elongated travel section does have the advantage of allowing some plug flow characteristics; for large basins, the potential for short-circuiting is greatly reduced. And if the aeration input zones are kept far apart (as would be the case if high oxygen input rates are kept per volume of passing liquid) the potential for anoxic zones and denitrification between the aerators are greatly enhanced. However, this same advantage can result in a disadvantage if the resultant effect is high in dissolved oxygen levels immediately downstream of the aerator. Intense oxygen input per volume of liquid passing through the aeration zone can result in aerator driving force inefficiencies.

The Carrousel is promoted by Envirotech; a sketch of the basin is shown on Figure 3. The oxygen transfer rate of the aerator is very good, being above 3.5 pounds of oxygen/hp-hr. Because of the vertical aerators location at the end of the channel and the partial loss of mixing expenditure, the mixing efficiency is most likely slightly less than the brush or disc.

The maintenance requirements of the low-speed aerator device used with the Carrousel is similar to that of other conventional low-speed vertical-shaft aerators.

## U-Tube Ditches

The U-Tube or draft-tube channel is promoted by Lightnin and is unique in that it does not provide an unrestricted flow circuit like that found with the other types of ditch systems. Oxygen delivery and mixing are independent with this system. A draft-tube turbine forces the channel flow through a conduit below the channel floor and behind a cross-sectional wall that acts as a dam across the channel width. In order to achieve sufficient channel velocities, the pumping rate of the turbine can easily be calculated; usually the rate would be 100 to 200 times the incoming flow rate. Air spargers located just below the impeller of the turbine results in fine bubbles being forced through the U-Tube. Usually there is only one air application point within the channel; this should enhance the possibility of anoxic zones at some point further downstream. Figure 4.

Oxygen transfer efficiencies should be good because of the increased pressures and forced routing through the U-Tube. Mixing efficiencies, however, might be poor because of the dam constriction across the width of the channel. One strong advantage with this system is the fact that all of the channel contents is forced through the U-Tube and must mix with the applied oxygen after each flow circuit. As of yet, however, there is no published data to indicate that passing the total

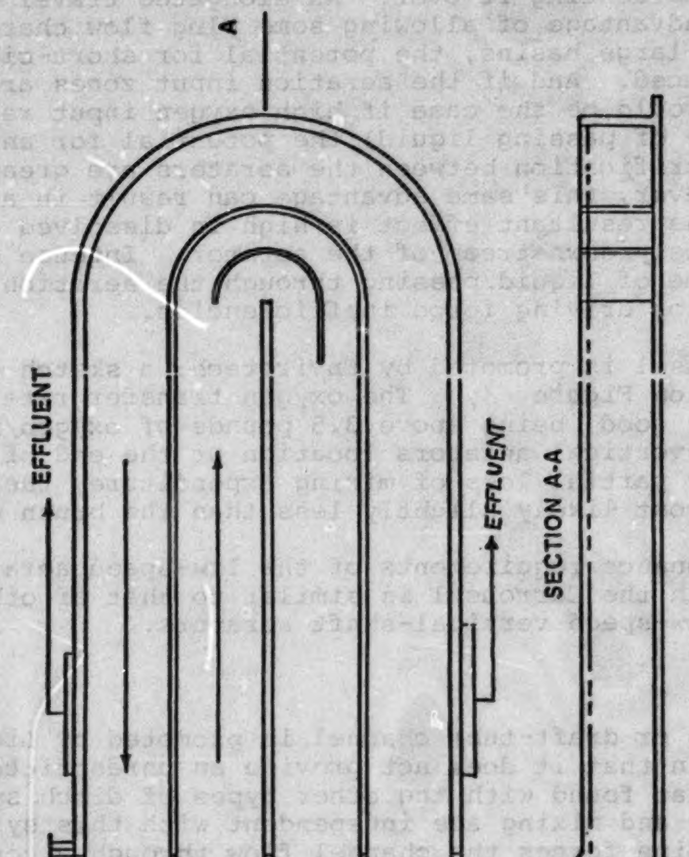
A
EFFLUENT
EFFLUENT
SECTION A-A
R.A.S.
INFLUENT
A

Figure 3. Carrousel Layout

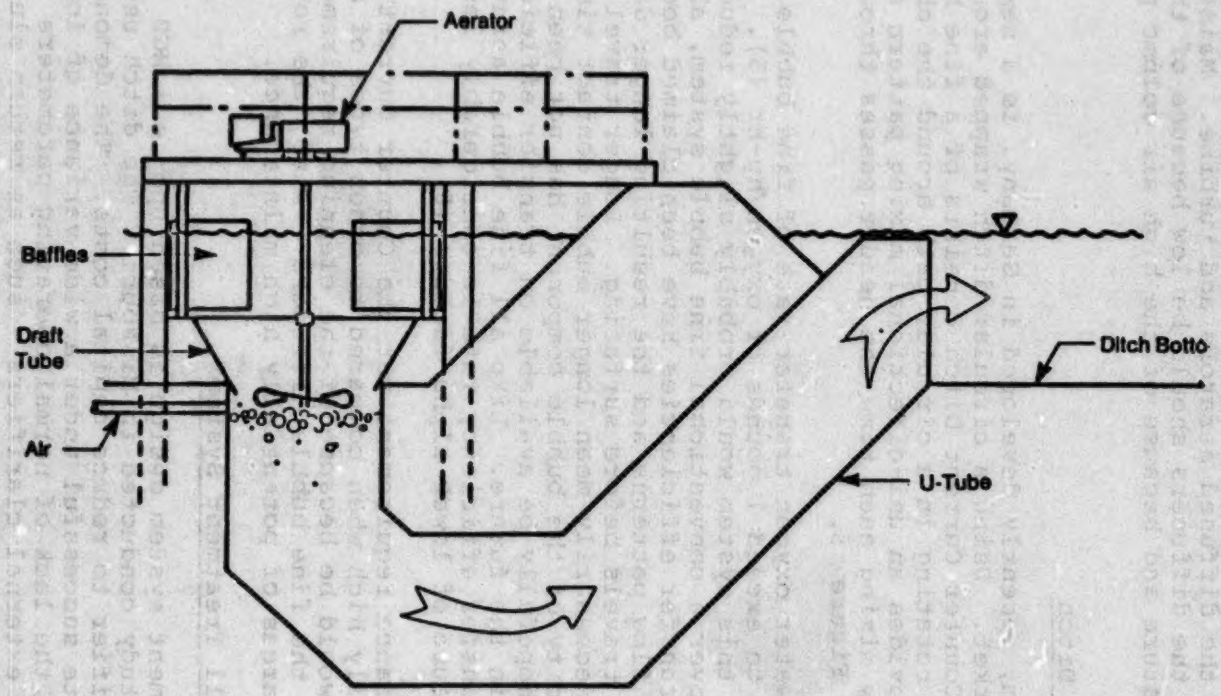
Aerator
Baffles
Draft Tube
Air
Ditch Botto
U-Tube

Figure 4. Aerator/U Tube - Longitudinal Section

contents of the basin through the U-Tube has extra process treatment benefits.

Maintenance requirements with the U-Tube ditch can be divided between the diffused aerators and turbine. Maintenance requirements on the diffusers should be low because of their coarse bubble nature and because of the high air volume per diffuser.

## Counter Current Ditch

This system, recently developed in Germany, is a newcomer on the ditch market. Using a circular ditch wrapped around a clarifier, the Counter Current Ditch consists of a fine bubble diffuser header rotating in a circular path around the channel. The rotation provides an uni-directional mixing pattern and localized upflow mixing each time the header passes through a different zone. Figure 5.

The clean water oxygen transfer rate of fine bubble systems have been shown to exceed 7 pounds of oxygen/hp-hr (5). The rotating arm of this system would probably slightly reduce the energy savings over a conventional fine bubble system, although higher oxygen transfer efficiencies have been claimed because of the directional flow patterns and the resulting longer distance that the bubble travels before surfacing. Longer travel distance does not necessarily mean longer bubble contact time if the upward velocity of the bubble component has not been altered. More data will hopefully be available on transfer efficiencies of this system in the future. Like all fine bubble aeration systems, the transfer efficiency rate is considerably less in dirty water because of lower alpha factors (6).

The maintenance requirement for the Counter Current Ditch can be potentially high when compared to other types of ditch systems. This would be because of the cleaning requirements associated with the fine bubble diffusers. Also, the rotating air header has areas of potentially high maintenance.

## Burns & McDonnell Treatment System

This treatment system design is based upon a 1 MGD demonstration study conducted in Missouri. The ditch used an in-channel clarifier to reduce capital costs. The demonstration program was quite successful under a wide variance of load conditions despite the lack of normal operating parameters associated with external clarifiers - such as return sludge pumping rates and sludge blanket level control. Figure 6.

One important development with this system is the unique combination of fine bubble diffusers with low speed submersible mixers. The mixer used in the program had an eight foot

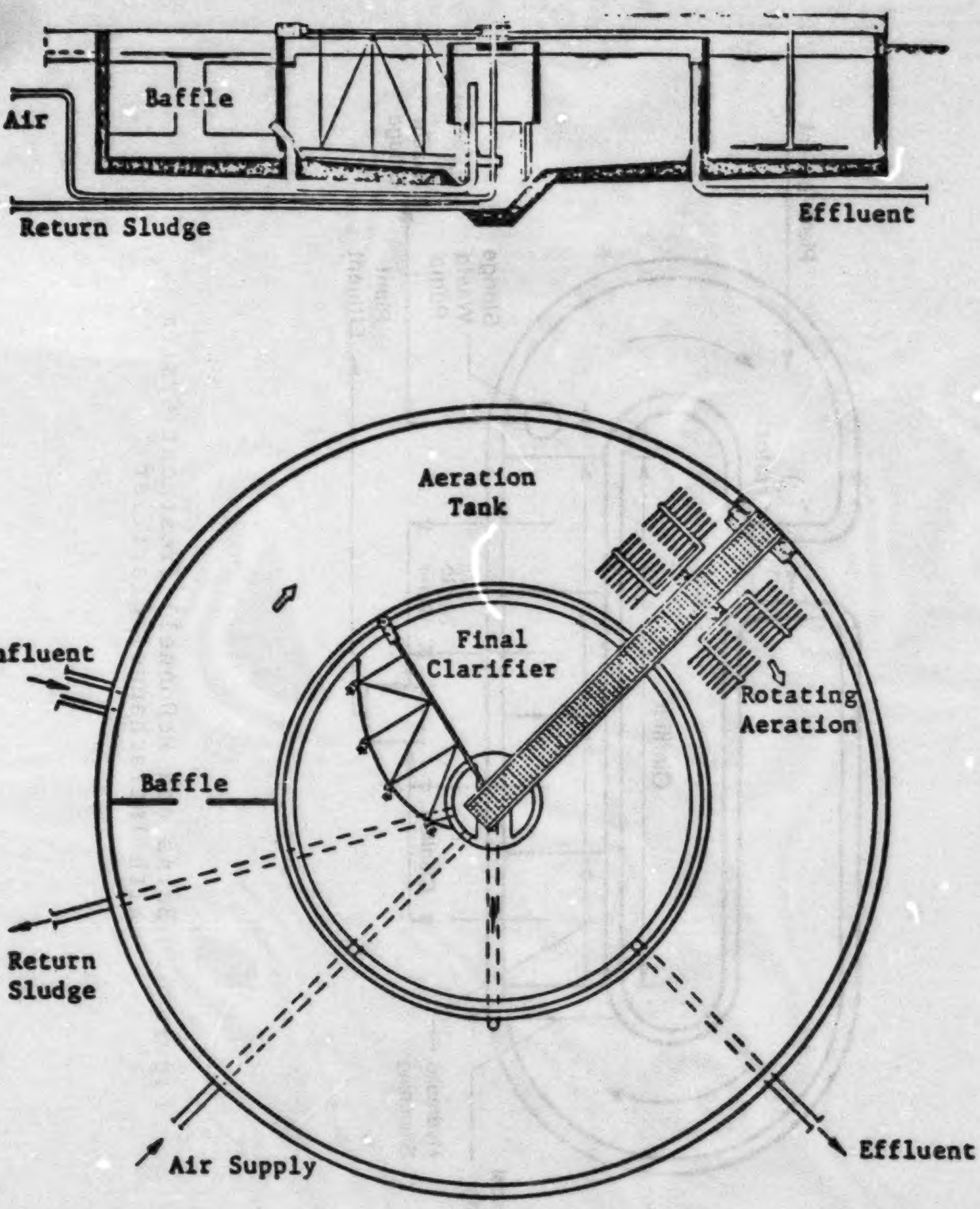
Baffle
Air
Return Sludge
Effluent
Aeration Tank
Final Clarifier
Influent
Rotating Aeration
Baffle
Return Sludge
Air Supply
Effluent

Figure 5. Counter Current Ditch

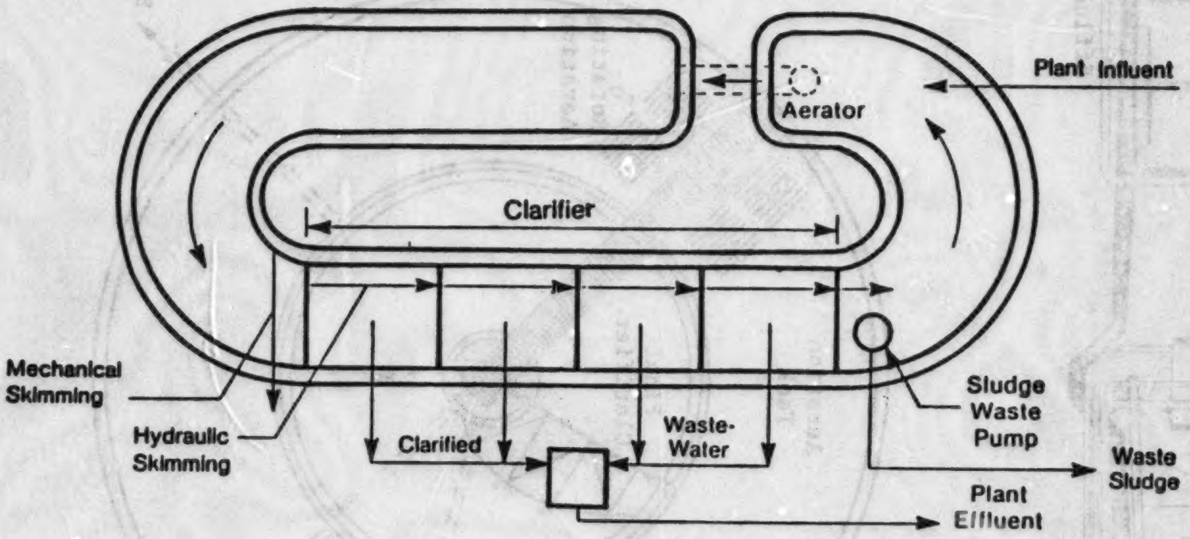
Plant Influent
Aerator
Clarifier
Mechanical Skimming
Hydraulic Skimming
Clarified
Waste-Water
Sludge Waste Pump
Waste Sludge
Plant Effluent

Figure 6. Burns and McDonnell Treatment System With Intrachannel Clarifier

diameter impeller and a 30 rpm rotation speed. The motor used was only 3½ hp for a 300,000 gallon basin; considering the constriction of the in-channel clarifier, the mixing efficiency of this device is very impressive. With the in-channel clarifier eliminated, the same 3½ hp mixer could be used for a 1,000,000 gallon basin and proper mixing would be provided.

Although the treatment system is very new, plants are now being designed by Burns & McDonnell for full scale operation.

## DISCUSSION

In discussing comparisons of the various systems, it is important to note that the types of equipment and basin configurations can overlap. Several single loop oxidation ditches operate using disc aerators, while there are some orbal multichannel plants operating using brush aerators. There is even one operating ditch plant in Georgia with both disc aerators and brush aerators in the same basin. And while all of the systems previously noted, except for the orbal were single compartment reactors, it is not difficult to imagine designs using two or more oxidation ditches or Carrousels or jet aeration channels in series, duplicating the process flexibility advantages of the orbal. In fact, such type designs have been used in Europe (7).

Many of the systems have particular areas of strength. The brush and disc ditches are more ideal for the smaller plants (though this by no means eliminates them from consideration for larger plants) where basin depths can be shallow and the oxygen input per aerator is small. The Carrousel is more practical for plants larger than 1 MGD; it would seem that plants larger than 50 MGD would have concrete intensity disadvantage because of the numerous folds to the basin. Jet aeration channels can be ideally used where deep basins are required.

The aerator types can be divided between surface types and sub-surface types. Sub-surface types will usually give the higher transfer rates in clean water; in dirty water, the transfer rates are not that far apart. If power costs associated with oxygen delivery is going to be an important consideration in the design, aerator flexibility becomes important for all of the systems. All of the looped reactors have the benefit of low power required for mixing - an important consideration when taking into account the low-load nature of the basin.

Maybe the most important area of energy savings is that of operating dissolved oxygen levels. Excessive levels will mean an energy loss. Near-zero levels are required to establish anoxic zones and oxygen recovery through denitrification. Running looped reactors in series can be an important energy saver; this can even be done with conventional low-loaded aeration tanks in series (7). The savings can be as high as 30 to 40% if the first reactor is kept near zero. If the last reactor is kept small in volume and high in dissolved oxygen, the process results will be maximized and the power costs optimized.

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## DISCLAIMER

The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency. and no official endorsement should be inferred.

# TRANSLATION OF CLEAN TO DIRTY WATER OXYGEN TRANSFER RATES

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## ABSTRACT

The translation of clean water oxygen transfer rates to field conditions represents a critical step in the design of aeration systems. This paper reviews the parameters required to make this translation, discusses common difficulties that arise and proposes methods that may facilitate more sound design decisions.

## INTRODUCTION

Whereas the aerator manufacturer reports the performance of aeration equipment as the oxygen transfer efficiency in clean water at 0 D.O., one atmosphere pressure, and 20°C, it is the transfer efficiency under actual process conditions that is of practical concern in the design of aeration systems. It is generally the responsibility of the design engineer to translate the manufacturer's clean water transfer rate to that expected under actual design conditions. This paper (a) addresses the fundamental aspects of oxygen transfer that should be recognized by the design engineer, (b) points out common misconceptions and difficulties that arise when translating clean water performance to field conditions and, (c) proposes methods which may facilitate more sound design decisions. Although geometry effects are important when relating shop test data to the actual aeration tank design, the problem of scale-up from one clean water system to another is not discussed in detail here. However, the influence of geometry on the determination of the field oxygen transfer rate relative to the clean water rate is examined. This paper is not written to provide correction factors to permit the quick and easy assessment of process performance of an aeration device or system. The evaluation of dirty water performance is a complicated task with many areas of uncertainty that still need to be resolved. Rather, the authors hope to shed some light on the variables that affect the performance of aeration equipment in the field and the manner in which those variables may be assessed.

## FACTORS AFFECTING DIRTY WATER PERFORMANCE

The standard clean water transfer rate (SOTR) as measured at 20°C with a dissolved oxygen concentration of zero may be related to actual field conditions ($OTR_f$) according to the following equations:

$$OTR_f = (SOTR)(\alpha)\left(\frac{\tau\Omega\beta C^*_{20} - C}{C^*_{20}}\right)(\theta^{T-20}) \qquad (1)$$

where $$\alpha = \frac{K_{la}\ (\text{field})}{K_{la}\ (\text{clean water})} \quad (2)$$

$$\beta = \frac{C^*\ (\text{field})}{C^*\ (\text{clean water})} \quad (3)$$

$$\tau = \frac{C^*_{sT}}{C^*_{s20}} \quad (4)$$

$$\Omega = \frac{P_b + \gamma d_e - P_{v20}}{P_s + \gamma d_e - P_{v20}} \quad (5)$$

and $$\theta^{T-20} = \frac{K_{la(T)}}{K_{la(20)}} \quad (6)$$

(Note: See appendix for definition of the terms used above.)

## Beta

The beta factor, sometimes referred to as the salinity correction, arises due to the "salting-out" effect of wastewater. That is, dissolved salts and organics tend to lower the saturation concentration of oxygen wastewater compared to that of distilled water. Although the activity of a saturated oxygen-water solution is by definition independent of dissolved contaminants, the concentration changes as the activity coefficient is altered by the salting-out effect.

The fact that the activity is independent of the purity of the water has important implications in the measurement of the saturation concentration under field conditions. Direct determination of the oxygen concentration in wastewater by the Winkler Method (1) is often complicated by oxidizing or reducing compounds which prohibit use of this method. Unfortunately a membrane covered probe theoretically responds to activity which is dependent on the degree of saturation - not the absolute concentration. Therefore a probe standardized in clean water will not yield a true reading of the dissolved oxygen concentration in a contaminated water solution.

To overcome these analytical difficulties, Bass and Shell (2) recommend applying a correction factor for temperature, barometric pressure, and dissolved solids concentration. Similarly, Standard Methods proposes an equation to adjust the saturation D.O. to field conditions (1).

## Theta

Since the overall oxygen transfer coefficient, $K_{la}$, is a function of liquid viscosity (among many other factors), a change in the liquid temperature which brings a concomitant change in viscosity will alter the transfer coefficient. Since very few aeration systems operate strictly at 20°C, a temperature correction in the form of equation (6) is commonly used.

Conflicting reports in the literature may be found concerning the value of theta. Temperature affects not only the physical properties of the gas-liquid interface, but also the diffusivity of oxygen in water. Since the gas-liquid interface is influenced in different ways by various aeration devices according their degree of turbulence, mixing patterns, bubble retention times and velocities, degree of droplet formation and spraying, it may be expected that temperature would affect each generic device to a varying degree thereby producing different theta values. Indeed, Hunter (3) reports that the water temperature correction factor depends on the level of turbulence and is, therefore, a function of the aerator. Hunter further cites several articles by other researchers to support his findings. Kalinske (4) claims that high energy systems will not exhibit temperature effects to the degree that low turbulence systems will and hence high turbulence aerators will have a lower value of $\theta$ than more quiescent aeration systems.

In contrast to those findings are the observations of Boon (5) and Gilbert (6). Boon states that there is approximately a 2% increase in $K_{la}$ per degree centigrade in the range from 0°C to 30°C for all types of aerators. Gilbert (6) presents data to show that the value of $\theta$ ranges between 1.024 and 1.026 which is roughly equivalent to a 2%/°C relationship. Gilbert did not specify the type of aerator used to generate this data.

Despite the reported variation of $\theta$ with aerator type, there is currently wide acceptance of a value of 1.024 for all aeration systems. Caution should be exercised by the engineer, however, when designing for conditions outside of the range from 10°C to 30°C. As the temperature approaches these extremes the theta value becomes critical.

## Alpha

The overall mass transfer coefficient, $K_{la}$, is often dramatically altered under field conditions due to dissolved and suspended contaminants. This ratio of the dirty water $K_{la}$ to the clean water $K_{la}$ is probably the most controversial and researched parameter that is used in the translation of clean water data to actual field conditions. Alpha factors as low as 0.2 (7) and as high as 1.5 (8) have been reported in the literature. Because of the extreme range of alpha, it will normally have the greatest effect on the translation of clean water to dirty water transfer rates. Variables affecting the value of alpha are numerous and complex and include aerator type, nature of the contaminants in wastewater, level of turbulence, mixing patterns, liquid depth, and air flow rate.

Since oxygen transfer is a mass transfer operation involving both the dissolution of a slightly soluble gas into a liquid as well as the transport of the oxygen throughout the bulk of the liquid, it is necessary to examine the effect of contaminants on both of these components of the mass transfer coefficient.

Basically, it may be stated that contaminants do not affect the bulk transport of oxygen (i.e. eddy diffusivity) to a great extent. Although some researchers (4) have pointed out that suspended solids and other compounds alter the pumping characteristics of an aeration tank, this effect is small compared to the impact of surfactants on the gas-liquid interface. Casey and Karmo (9) and Poon and Campbell (10) have indicated that suspended solids, especially bio-solids, alter the clean water transfer coefficient, but this effect is relatively minor over the range of mixed liquor solids maintained in typical activated sludge systems. In contrast, surfactants can exert a significant influence on the gas-liquid interface by virtue of their hydrophobic end which is attracted to the gas phase while the hydrophilic head remains drawn to the liquid. This surface activity decreases the surface tension of the interface which can result in the formation of smaller bubbles in a diffused air system, and the production of smaller droplets and greater gas entrainment in surface aeration systems. More importantly, these surfactants alter the hydrodynamic activity of the gas-liquid interface. Mancy and Barlage (11) have claimed that surface active agents form a highly viscous layer (on the order of $10^4$ poise; approximately that of butter) which causes a greater path length for oxygen in the liquid film surrounding an air bubble (or liquid droplet). This viscous layer also reduces the rate of shear around the bubble which lowers the rate of surface renewal (12).

This surface effect will influence the transfer rate of different types of aerators in vastly different ways. Holroyd and Parker (13) found that turbulent surfaces are less susceptible to surface active agents than are more quiescent surfaces. Kessener and Ribbius (7) reported alpha values of 0.2 for diffused air systems compared to 0.82 for brush aerators. Systems that continuously form fresh air-liquid interfaces through violent mixing are usually not adversely affected by surfactants. In fact, the formation of smaller droplets along with greater bubble entrainment in a mechanical aeration system can often result in alpha values greater than unity (8). On the other hand, oxygen transfer devices that promote transfer from the gas phase to the liquid phase by virtue of a relatively gentle gas-liquid mixing action may allow surfactants to accumulate at the interface with deleterious results.

Thus, surface aerators would not be expected to experience the substantially depressed alpha values that are frequently measured for fine bubble diffusers. At the same time, submerged diffusers may be affected to varying degrees by surface active agents. As previously mentioned, oxygen transfer consists of gas to liquid transfer followed by transport throughout the bulk liquid. An aeration tank with a non-uniform diffuser arrangement such as a spiral-roll coarse bubble system may be divided into an aeration zone (in the immediate vicinity of the diffusers) and a pumping, or mixing zone. Since surfactants have only a minor effect on the transport of oxygen in the pumping zone, the reduction of alpha will depend on the relative contribution of the aeration zone to $K_{la}$. For aerators where the mixing component of $K_{la}$ is a significant factor, the alpha value may be close to unity. However, it should not be assumed that a coarse bubble aerator will necessarily be characterized by an alpha value greater than a fine bubble aerator. Downing and Bayley (14) have measured similar values of alpha for both large and small bubbles rising in a narrow column. The degree of bulk mixing and the eddy diffusivity of oxygen are important determining factors of alpha, not just the bubble size.

The design engineer should also recognize that alpha is subject to variation. This variation may occur on a seasonal, daily, or even hourly basis. Gilbert (6) has noted a variation in alpha from 0.8 to 0.5 from hour to hour in a brewing wastewater. Eckenfelder (15) has also shown a similar variation for Kraft mill wastes. While industrial waste variations may be particularly extreme due to fluctuations in wastewater strength and composition, it is also possible for municipal treatment plants to experience variation in alpha values. However, the nature of the wastewater (i.e. industrial contributions, etc.) will make the degree of variability site specific. Doyle (16) conducted tests in a 10 foot deep, 1 foot diameter column using a fine bubble ceramic diffuser and found little daily or hourly variation in Milwaukee primary effluent. Alpha values were measured in the range of between 0.45 to 0.55 (see Figure 1).

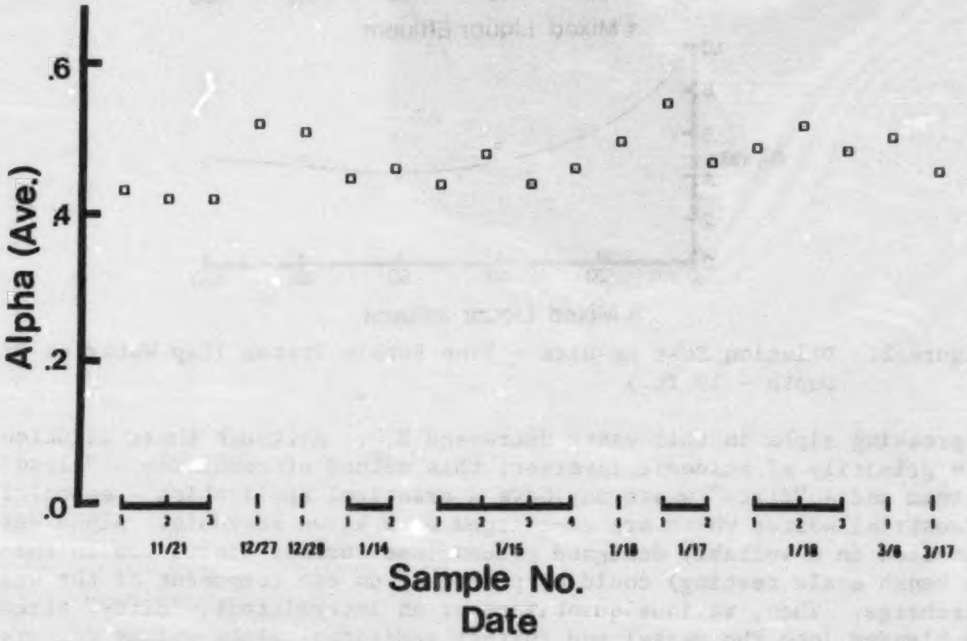

Figure 1. Fine Bubble System Alpha Measurements - Mixed Liquor Influent; Average of Replications (Depth - 10 ft.)

These tests were performed at the Milwaukee Jones Island wastewater treatment plant which receives a heavy industrial load and, hence, conclusions relative to the variability of alpha for strictly domestic wastewater cannot be made.

An explanation for the relative constancy of alpha at this plant was provided by conducting dilution tests in which the mixed liquor feed was progressively diluted with tap water. As Figure 2 reveals, even small amounts of this wastewater added to tap water will lower alpha substantially. Apparently even low concentrations of the compound(s) responsible for

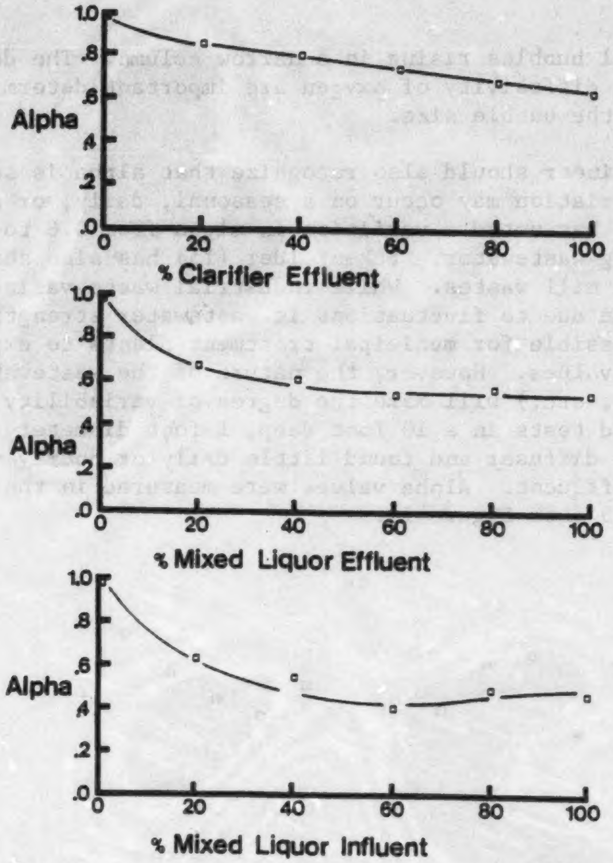

Figure 2. Dilution Test Results - Fine Bubble System (Tap Water as Diluent, Depth - 10 ft.)

depressing alpha in this waste decreased $K_{la}$. Although these dilution tests are primarily of academic interest, this method of combining a "clean" stream and a "dirty" waste may have a practical application - especially for industrial wastes which are discharged on a known schedule. Alpha determination in a suitably designed column (see further discussion in this paper on bench scale testing) could be performed on one component of the waste discharge. Then, various quantities of an intermittent, "dirty" stream could be blended into the vessel and further additional alpha values measured. If the discharge volume and duration is known for the "dirty" stream, a design estimation of alpha could be made.

In addition to temporal variation, alpha has been found to change during the course of treatment. Lister and Boon (17) show a recovery in alpha as the biodegradable agents responsible for lowering the field $K_{la}$ are removed. The linear relationship shown in Figure 3 is intended only to illustrate a concept. The increase in alpha with treatment may not be linear and will depend not only on the treatment kinetics and tank hydraulics, but also on the biodegradability of the contaminants. Doyle (16) has noted a general increase in alpha with treatment as shown by Figure 4. These successive batch tests were made with a single sample drawn from the mixed

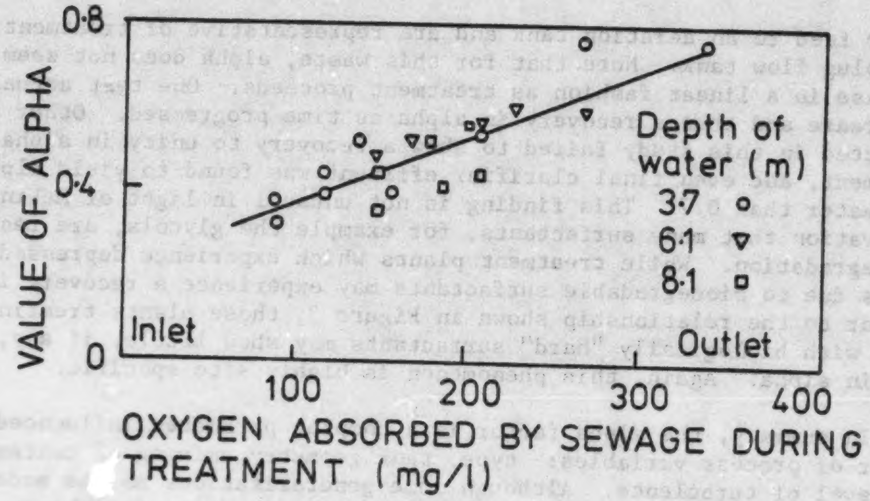

**Figure 3. Variation of Alpha with Degree of Treatment (After Lister and Boon (17)).**

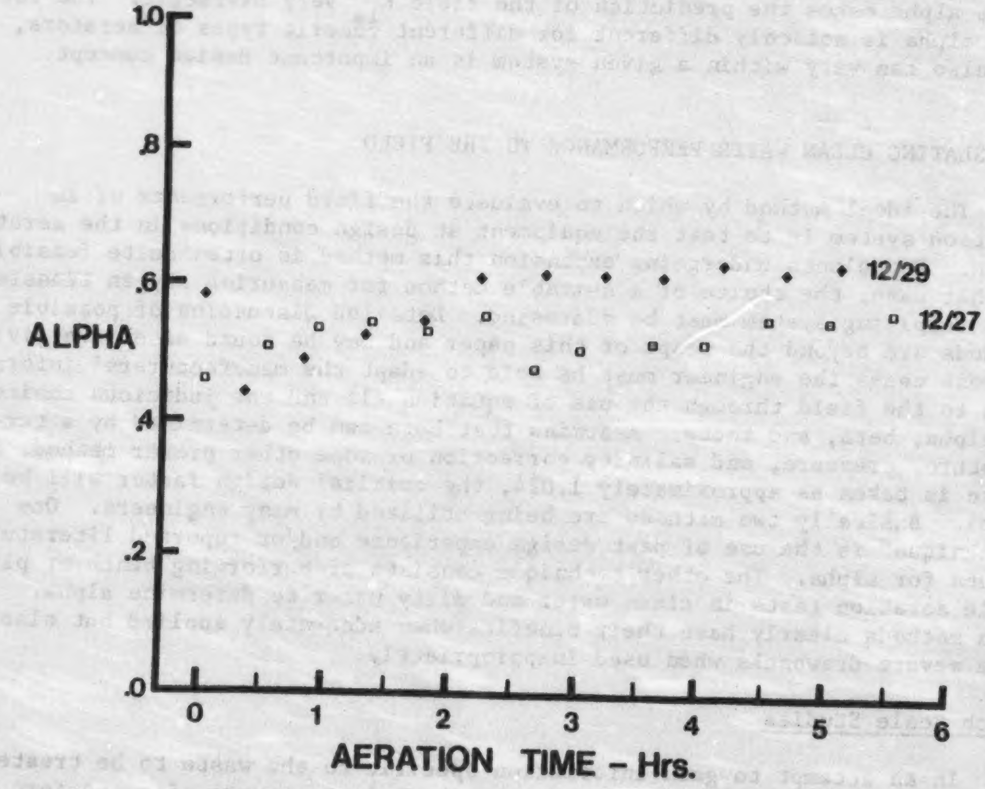

**Figure 4. Effect of Treatment on Alpha - Batch Tests (Depth - 10 ft.)**

liquor feed to an aeration tank and are representative of treatment in a perfect plug flow tank. Note that for this waste, alpha does not seem to increase in a linear fashion as treatment proceeds. One test actually showed a decrease and then a recovery in alpha as time progressed. Other data collected in this study failed to show a recovery to unity in alpha with treatment, and even final clarifier effluent was found to yield alpha values no greater than 0.7. This finding is not unusual in light of McKinney's (18) observation that many surfactants, for example the glycols, are resistant to bio-degradation. While treatment plants which experience depressed alpha values due to biodegradable surfactants may experience a recovery in alpha similar to the relationship shown in Figure 3, those plants treating wastewater with biologically "hard" surfactants may show little, if any, improvement in alpha. Again, this phenomenon is highly site specific.

In summary, the alpha factor is a complex parameter influenced by a number of process variables: type, tank geometry, mixing of contaminant, and level of turbulence. Although some generalizations may be made regarding the effect of contaminants on alpha, the lack of a unifying theory for these variables and a paucity of knowledge concerning the mechanism that determines alpha makes the prediction of the field $K_{la}$ very difficult. The fact that alpha is not only different for different generic types of aerators, but also can vary within a given system is an important design concept.

## TRANSLATING CLEAN WATER PERFORMANCE TO THE FIELD

The ideal method by which to evaluate the field performance of an aeration system is to test the equipment at design conditions in the aeration basin. For plants undergoing expansion this method is often quite feasible. In that case, the choice of a suitable method for measuring oxygen transfer in a respiring system must be addressed. Detailed discussion of possible methods are beyond the scope of this paper and may be found elsewhere (19). In most cases the engineer must be able to adapt the manufacturers' information to the field through the use of equation (1) and the judicious choice of alpha, beta, and theta. Assuming that beta can be determined by a temperature, pressure, and salinity correction or some other proper method, and theta is taken as approximately 1.024, the critical design factor will be alpha. Basically two methods are being utilized by many engineers. One "technique" is the use of past design experience and/or reported literature values for alpha. The other technique consists of performing bench or pilot scale aeration tests in clean water and dirty water to determine alpha. Both methods clearly have their benefits when adequately applied but also have severe drawbacks when used inappropriately.

### Bench Scale Studies

In an attempt to gain information specific to the waste to be treated, especially unique industrial wastes, bench scale tests are often performed. Various authors (2, 20, 21) have described bench scale alpha units. A bench test is useful only if it can be translated to full scale operation. Unfortunately this is a difficult undertaking due to the extremely complex components of mass transfer. Accordingly, the literature is replete with reported failures to duplicate the full scale alpha on a small scale. Barnhart (22), Otoski et al. (23), Schmit et al. (24), and Gilbert (6) have all experienced an inability to match alpha on the large and small scale. While Stukenberg (20) states that equilibrating the clean water oxygen transfer coefficient in the small vessel and large tank will yield similar alpha values during dirty water testing, it must be realized that most researchers have found that greater air/volume ratios must be used in the small scale units due to decreased transfer efficiency at lower liquid depths (for diffused air systems, see Figure 5). Dissimilar air/volume ratios can cause disparate mixing intensities which will result in different determinations of alpha. Mixing, in general, is a difficult property to scale, especially when attempting to maintain similitude with other aspects of mass transfer. The scaling of a surface aerator is complicated by variables such as bubble entrainment and drop size. Shell (25) has stated that a mechanical aerator is impossible to scale correctly. It is fortuitous that the determination of alpha for a mechanical aerator is not highly critical since the value is fairly close to unity.

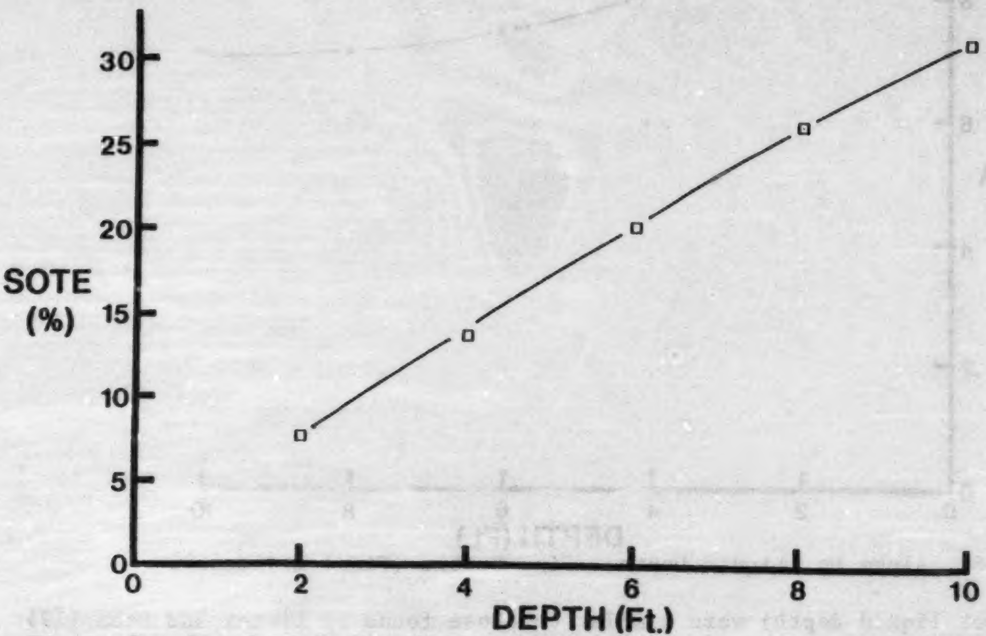

Figure 5. Standard Oxygen Transfer Efficiency vs. Depth - Fine Bubble (Air Flow - 0.5 SCFM/Diffuser).

Although simply matching $K_{la}$ on the large and small scales does not appear to be an adequate criteria for the design of a bench scale model, and although mechanical mixing may not lend itself to modeling, the design of a small scale vessel for certain aerators appears to be feasible. For a fine bubble diffuser system employing uniform floor coverage, the diffusers have often been observed to act as separate mixing cells (26). This type of mixing may lend itself to modeling in a tall, narrow column. Field conditions such as type of diffuser, air flow rate, percent floor coverage of the stones, and liquid depth should be duplicated as closely as possible. The use of a shallow test vessel can produce misleading estimates of alpha. Doyle (16) found that alpha was a function of depth in studies conducted in a one foot diameter column using a ceramic fine bubble diffuser (See Figure 6). Values of alpha (0.45-0.55) in the mixed liquor feed (measured at a 10 foot liquid depth) were similar to those found by Lister and Boon (17) in full scale tests. Although it may be desirable to operate at a liquid depth even closer to the actual aeration tank depth, clearly a tall column is an improvement on the apparatus typically recommended in the literature.

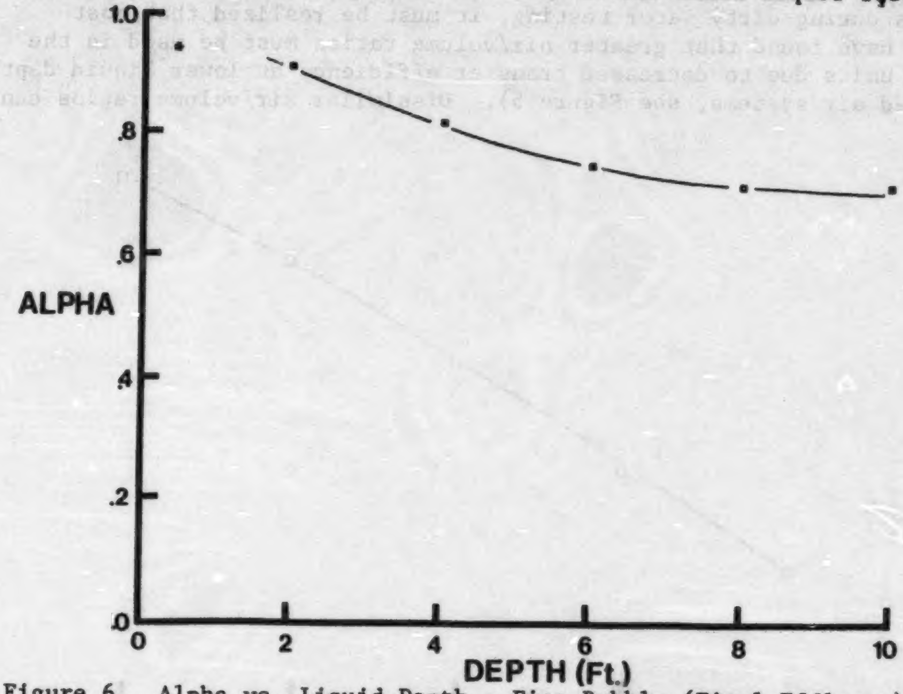

Figure 6. Alpha vs. Liquid Depth - Fine Bubble (Final Effluent).

A tall, narrow column will not be a suitable vessel for modeling all types of diffused air systems, however. Morgan and Bewtra (27) stated that a spiral-roll, coarse bubble aerator cannot be successfully modeled in a narrow column. Results from Doyle's work substantiate this finding. A coarse bubble aerator was simulated by installing a plate with two 1/16" holes in the bottom of the column. A value of alpha (0.4-0.5) equivalent to that of the fine bubble system resulted. Since bulk mixing was largely eliminated by the restrictions of a narrow column, this component of $K_{la}$ which is largely unaffected by surfactants was not incorporated in the column tests. Hence, alpha was probably lowered below levels that would have been observed had this mixing component been correctly modeled.

To summarize, the bench scale apparatus typically recommended in the literature for the determination of alpha has often been found to be inadequate. However, there are certain applications for small scale test vessels which can be used to model some types of aeration systems. In particular, a tall, narrow column utilizing an actual commercial diffuser and operated at close to the design conditions of liquid depth and air flow rate has been found to be useful in assessing alpha for a fine bubble diffuser. Attempts to model aeration systems where mixing and liquid pumping comprises a significant portion of $K_{la}$ are more complex. Proper scaling of these properties is vital when attempting to evaluate alpha on the small scale. Further research into this type of modeling is necessary if small scale estimates of alpha for these aeration systems are to be made with reliability.

## Engineering Estimation

The use of past experience and/or reported literature values of alpha, a technique referred to herein as "engineering estimation", is probably the most common method currently employed for selecting a design value for alpha. A thorough review of the literature indicates that there is very little information available on actual field measurements of alpha in any type of wastewater. It is likely that a significant number of engineers use alpha values quoted in standard textbooks or journal articles. One of the chief references often cited is Eckenfelder (15). Eckenfelder, in 1959, listed alpha values for many industrial wastes as well as municipal wastes, these results all gathered from tests in a 1 liter (or 4 liter) vessel. There is no evidence in the literature that these values have ever been corroborated.

Unfortunately, the verification of alpha values collected from the laboratory is most difficult to achieve since methods used to estimate field transfer efficiency in dirty water are often imprecise and inaccurate. Furthermore, in order to accurately assess alpha one must also have field clean water transfer data as well. Probably one of the earliest field tests for alpha was performed by Kessener and Ribbius (7) in 1933. They found alpha values of 0.2 for a diffused air system and 0.8 for a brush aerator, both in sterilized wastewater. Since that time, little work was reported until the Water Pollution Research Laboratory in Stevenage, England began extensive testing in large tanks in the late 1950's. Much of this work was done with anionic detergents since it was felt that the surfactant in wastewater played the most important part in depressing oxygen transfer rates. Parallel tests were also performed with mixed liquor. Since that time, numerous other investigators have tested a variety of aeration devices in large tanks employing anionic detergents (17, 28, 29, 30, 31). A summary of much of this data is presented in Table 1. It should be emphasized that the values of alpha obtained from detergent tests may give an indication of the depression of the oxygen transfer rate that will be observed in wastewater, but they should not be considered as absolute.

Field testing in dirty water, as mentioned earlier, is a far more complex problem than detergent testing. The biological uptake rate in the wastewater creates an analytical problem which may result in significant error (19). Testing with or without mixed liquor solids adds to the difficulty. Finally, spatial and temporal wastewater variability creates further uncertainty as to the true value of alpha to use for design purposes.

Results of large scale tests for alpha in dirty water appear in Table 2. Steady state testing by measuring oxygen uptake rates in mixed liquor during

Table 1. Alpha Values Determined in Large Systems Employing Detergents*

| Aerator Type | Tank Size (gal) | Characteristics | Alpha | Reference |
|---|---|---|---|---|
| Fine Bubble | 48000 | Domes, Subm - 20 ft. | 0.48 | Lister & Boon (17) |
| Fine Bubble | 48000 | Domes, Subm - 27 ft. | 0.37 | Lister & Boon (17) |
| Fine Bubble | 39000 | Saran & Brandol tubes; Plastic box; Subm $\cong$ 12 ft. | 0.6 | von der Emde (28) |
| Medium/Coarse Bubble | 39000 | Perforated Tubes Subm ~ 13 ft. | 0.7-0.8 | von der Emde (28) |
| Coarse Bubble | 35000 | Air Flow-25 scfm/1000 ft$^3$; Subm - 20 ft. | 0.65 | Schmit et al. (24) |
| Coarse Bubble | 35000 | Air Flow-90 scfm/1000 ft$^3$; Subm - 20 ft. | 0.75 | Schmit et al. (24) |
| Coarse Bubble | 87000 | Air Flow - 80-190 scfm/1000 ft$^3$; Subm - 10 ft. | 0.7-0.8 | Otoski (23) |
| Cage Rotor | 50000 | - | 1.2 | Downing et al. (29) |
| Simplex Aerator | 500000 | Compared against Secondary Effluent | > 1.1 | Downing et al. (29) |
| Simplex Cone | Full Scale | - | 1.1 | von der Emde (28) |
| Kessner Brush | Full Scale | - | 1.0-2.0 | Baars (30) |
| Surface Aerator | Full Scale | - | 0.93 | Boon (31) |
| Static Tubes | 87000 | Gs 10-190 scfm/1000 ft$^3$ | 1.1 | Otoski (23) |

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[*] All tests with tap water and detergent (5 mg/l); oxygen transfer determinations by unsteady state procedure.

Table 2. Alpha Values Determined in Large Systems Employing Wastewater

| Aerator Type | Test Procedures | | Tank Size (gal) | Characteristics | Alpha | Reference |
|---|---|---|---|---|---|---|
| | Clean Water | Dirty Water | | | | |
| Fine Bubble (Domes) | NSS | SS/OG | 4800 | Mixed Liquor, inlet, Subm - 20 ft. | 0.49-0.56 | Lister & Boon (17) |
| Fine Bubble (Domes) | NSS | SS/OG | 4800 | Mixed Liquor, inlet, Subm - 27 ft. | 0.43-0.49 | Lister & Boon (17) |
| Fine Bubble (Discs) | NSS | Tracer | Full Scale | Mixed Liquor | 0.4-0.5 | Kiiskinen (35) |
| Fine Bubble (Tubes) | ? | ? | 39000 | Final Effluent Subm - 13 ft. | 0.6 | von der Emde (28) |
| Fine Bubble (Domes) | Calc. | O.G. | Full Scale | Mixed Liquor Subm - 15 ft. | 0.42-0.98 (inf → eff) | Redmon et al. (34) |
| Fine Bubble (Discs) | Calc. | O.G. | Full Scale | Mixed Liquor | 0.34-0.62 (inf → eff) | Redmon et al. (34) |
| Fine Bubble (Tubes) | Calc. | O.G. | Full Scale | Mixed Liquor Cross roll | 0.46-0.85 (inf → eff) | Redmon et al. (34) |
| Coarse Bubble | Calc. | O.G. | Full Scale | Mixed Liquor Cross roll | 0.83-1.19 (Random) | Redmon et al. (34) |
| Coarse Bubble | Tracer | Tracer | 660,000 | Chemical Waste; Mixed Liquor; Floor grid; Subm - 19 ft. | 0.6-0.65 | Campbell (19) |
| Oxidation Ditch (Mech. Aer.) | NSS | SS | 194,000 | Mixed Liquor (Extd Aer) Subm - 1.5 ft. | 0.95-1.05 | Huibregtse & Doyle (32) |
| Oxidation Ditch (Mech. Aer.) | NSS | SS | 179,000 | Mixed Liquor; Refinery; Disc Subm - 1.75 ft. | 0.98-1.03 | Applegate & Huibregtse (33) |

NSS - Non steady state; SS - steady state; O.G. - off gas.

steady state conditions was performed by Lister and Boon (17), Huibregste and Doyle (32) and Applegate and Huibregste (33). Off gas analysis whereby inlet and outlet oxygen and $CO_2$ are analyzed was employed by Redmon et al. (34), Campbell (19), and Lister and Boon (17). The radioactive tracers Krypton-85 and tritiated water were used in field tests reported by Kiiskinen (35) and Campbell (19).

From the standpoint of accuracy and precision of testing under dirty water conditions, it is felt that tracer techniques and off gas methods offer great potential. Recent work with a dual non-steady state procedure employing hydrogen peroxide may also be effective (36, 37). Since laboratory scale tests have very limited usefulness in estimating dirty water alpha, field dirty water testing (following clean water tests) with a variety of aeration devices and wastewaters will have to supply the data base needed to provide the field with useful alpha values.

## SUMMARY

It is clear that the state of the practice in translation of clean to dirty water oxygen transfer rates is still in its infancy. With the escalation of energy costs, the need for efficient aeration systems is apparent. However, aeration system design appears to be lagging behind aeration device technology. Many new questions are being raised regarding the design and operation of aeration systems in light of changing constraints on that design. Conservative over-design is not necessarily the answer since it may lead to serious energy consuming inefficiencies. It will be necessary to better refine design practice and specifications so that designs more closely fit field situations. This means that the profession must follow up on designs with careful field evaluation. This is not a new concept - Conway and Kumke (21) made a similar plea in 1966 to the profession. Improvements in testing of oxygen transfer devices under clean and dirty water conditions will go a long way in providing the tools necessary to accomplish a part of this job.

## APPENDIX

$C^*_{20}$ - dissolved oxygen saturation concentration at 20°C at given submergence, $m/L^3$.

$C^*$ - dissolved oxygen saturation concentration, $m/L^3$.

$C^*_s$ - dissolved oxygen saturation concentration at 1.0 atmosphere and 100% relative humidity, $m/L^3$.

$C$ - effective average dissolved oxygen concentration in process liquid, $m/L^3$.

$K_{la}$ - overall liquid phase mass transfer coefficient for oxygen in water, $1/t$.

$d_e$ - effective saturation depth at infinite time, L.

$P_b$ - atmospheric pressure at field condition, $f/L^2$.

$P_s$ - standard atmospheric pressure, usually 1.0 atmosphere, at 100% relative humidity, $f/L^2$.

$P_v$ - saturated vapor pressure of water, $f/L^2$.
$T$ - temperature, °C.
$OTR_f$ - oxygen transfer rate under field conditions, M/T.
SOTR - standard oxygen transfer rate, M/T.
$\gamma$ - weight density, $M/L^3$.

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## AERATION SYSTEM SCALE-UP

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## ABSTRACT

The approach to aeration system scale-up and the degree of scale-up success for different types of aeration systems is discussed. Scale-up examples for surface turbines, cage rotors, diffusers, spargers, static systems as well as RBC system aeration are presented and discussed.

Progress in the area of aeration system scaling has been made, Predictions of oxygen transfer in full scale systems based on models developed from laboratory scale data have met with varying degrees of success.

## INTRODUCTION

Scale-up formulations for full-scale process design from data generated by way of prototype plant simulation at bench-scale have been the topic of many investigations. This is particularly true in the design of a variety of chemical engineering processes. Advances in this direction in the wastewater treatment process design field are few. Aside from the obvious benefits of having a reliable scale-up transform for a particular process, a further and perhaps even more important application would be the ability of scaling down process difficulties encountered in prototype plant operation and replicate the field condition in the laboratory for further study under controlled conditions.

Since this workshop concerns aeration, it would be appropriate to direct the focus of this presentation to the problems of and directions various attempts of devising scale-up functions for oxygen transfer into water have taken. By addressing the problem of oxygen transfer into water, and even if we are successful in scaling this process up or down, only one half of the problem is solved. For aerobic biological treatment to be successful the oxygen must also be transported to the cell surface for diffusion into the cell. There appears to be agreement that for biochemical reactions this represents the rate limiting step, provided of course that adequate substrate and nutrients are available. The transfer of mass into and out of cells can be enhanced, up to a limiting value, through the creation of turbulence.

## THE ROLE OF TURBULENCE

It is this turbulence phenomenon which results not only in the transfer of oxygen into cells but keeps the mixed liquor solids in suspension. We know that the energy input through an aeration device into a liquid system is translated into liquid movement, and that beyond a certain minimum energy input turbulent flow results. One manifestation of turbulence is eddies which come in all sizes.

In considering turbulence, two aspects are dominant...the scale of turbulence and the intensity of turbulence. The scale of turbulence refers to the mean size of eddies, while intensity of turbulence to the speed of angular rotation of the eddies.

In models, the scale of turbulence is small and the intensity of turbulence is high. For prototype systems, the opposite is true. Most of the energy imparted to a system through an aeration device will result in large eddies. They produce the main flow in the aeration tank. But large eddies have a geometric limitation. For instance they can not be larger than the liquid depth in the aeration tank. Ultimately eddies are dissipated. Thus large eddies must become smaller and smaller until finally they become extinct. It would stand to reason then that a whole continuum of eddy sizes between these two extremes exists. The dissipation of this energy is through friction (heat) caused by the general break-up of eddies due to collisions. Of greatest importance in biochemical processes are the smallest eddies which are capable of reaching the cells and thus are instrumental in the more rapid diffusion of oxygen, nutrients, substrate into and metabolic waste products out of the cells. The total turbulence energy is distributed among the large and these small eddies.

Requirements for successful process scaling concern meeting a variety of similarity criteria between model and prototype namely geometric, dynamic, kinematic and environmental similarity. One requirement for equal process results in small and large sized tanks would be geometric similarity in both turbulence scale and intensity. This similarity should of course extend to oxygen bubble size, the microorganisms and the floc particles. However, we are unable to scale or control the geometry of any of these.

Having recognized the importance of turbulence in mass transfer processes and the difficulties associated with its geometric similarity we are then faced with our greatest challenge in process scaling. How can we ensure that when examining a process at different scales the same turbulence level is retained? Is it possible to scale turbulence? If the answer is no, can these differences in turbulence levels between different sized systems and their concomittant effects on process results be accounted for in some other way?

It would appear reasonable that by involving a process response parameter, such as the overall oxygen transfer coefficient, $K_La$, successful process scaling might be possible.

If environmental and biological conditions can be kept identical, one should be able to scale bio-hydraulic systems provided mass transfer rates as expressed by the overall oxygen transfer coefficient ($K_La$) in both model and prototype are equal. Equal process results should be attainable.

## $K_La$ AS A SCALING PARAMETER

The overall oxygen transfer coefficient, $K_La$, is dependent on a variety of variables, and correlations of variables affecting $K_La$ have been presented (1-11). The number of variables influencing the magnitude of $K_La$ are also dependent on the type of aeration device. In recent years a number of researchers have made an attempt at using $K_La$ in a process scale up function.

A brief review of approaches to scale-up for surface aeration devices, diffused aerators, static aerators and cage rotors is presented. As well, considerations for scale-up of rotating biological contactors are highlighted. Fundamental to any investigation of this nature is an identification of operating and system variables for the various aeration system alternatives.

## SURFACE AERATION

### Scale-Up Function

It would be well then, to start by looking at a surface aeration system, identify and examine all possible variables which influence $K_La$, and then go on to the other aeration system alternatives. It will become apparent that a number of justified, simplifying assumptions can be made.

Variables influencing fluid motion in an unbaffled, surface turbine-agitated square tank can be identified as:

$$fn\ (W, D, HI, DI, WI, LI, B, S, \rho, \nu, g, N, P) = 0 \qquad (1)$$

For oxygen transfer, the corresponding influencing variables are:

$$fn\ (KLA, DI, HI, N, \nu, D_L, g, \rho, \sigma, P) = 0 \qquad (2)$$

In order to arrive at a general relation between variables influencing $K_La$ in a surface aerated turbulent flow regime, variables associated with fluid motion and mass transfer were combined into one functional relationship:

$$fn\ (W, D, HI, DI, WI, LI, B, S, \rho, \nu, g, N, P, KLA, D_L, \sigma) = 0 \qquad (3)$$

where:

$W$ = a significant tank dimension (L)
$D$ = liquid depth (L)
$HI$ = impeller immersion in liquid (L)
$DI$ = impeller diameter (L)
$WI$ = impeller blade width (L)
$LI$ = impeller blade length (L)
$B$ = number of impeller blades dimensionless
$S$ = impeller pitch (L)
$\rho$ = fluid density $(FL^{-4}T^{2})$
$\nu$ = fluid kinematic viscosity $(L^{2}T^{-1})$
$g$ = gravitational constant $(LT^{-2})$

$N$ = impeller speed $(T^{-1})$

· $KLA = K_La$ = overall oxygen transfer coefficient $(T^{-1})$

$P$ = power input $(LFT^{-1})$

$D_L$ = molecular diffusivity of oxygen $(L^2T^{-1})$

$\sigma$ = surface tension $(FL^{-1})$

Using dimensional analysis and selecting a length DI, a velocity N and fluid density $\rho$ as three of the independent variables and letting each remaining variable be one of the dependent variables, the sixteen variables of equation (3) were resolved into thirteen dimensionless parameters:

$$\text{fn}\left(\underset{1}{\frac{HI}{DI}}, \underset{2}{\frac{W}{DI}}, \underset{3}{\frac{W}{D}}, \underset{4}{\frac{WI}{DI}}, \underset{5}{\frac{LI}{DI}}, \underset{6}{\frac{S}{DI}}, \underset{7}{B}, \underset{8}{N_{Re}}, \underset{9}{N_{Fr}}, \underset{10}{N_P}, \underset{11}{\frac{KLA}{N}}, \underset{12}{N_{Sc}}, \underset{13}{N_{We}}\right) = 0 \qquad (4)$$

Equation (4) incorporates all variables describing a surface aerated system. Term 1 (HI/DI) is designated as an impeller immersion simplex which defines a depth of impeller immersion in the liquid. Terms 2 and 3 (W/DI, W/D) are terms defining the boundaries of the physical system which are usually fixed in the design of an aeration tank. Terms 4, 5, 6 and 7 (WI/DI, LI/DI, S/DI, B) describe the impeller geometry, terms 8 and 9 ($N_{Re}$, $N_{Fr}$) the influence of viscous and gravitational forces on the fluid system, while term 10 ($N_p$) is generally described as a parameter characterizing the general flow pattern of the fluid system. Term 11 ($K_La/N$), is defined as an oxygen transfer number and terms 12 and 13, the Schmidt ($N_{Sc}$) and Weber ($N_{We}$) numbers, characterizing the fluid system's physical properties and surface tension forces, respectively.

If geometric similarity between the various aeration devices and tanks is enforced, then the simplexes describing reactor geometry (W/DI and W/D) and the turbine impeller geometry (W/DI, LI/DI, S/DI, B) become constants in the scale-up function. If the fluid systems under consideration have the same physical properties, then the Schmidt number ($N_{Sc}$) which defines these properties also remains constant. Thus, in these systems, it is seen that the Reynolds number, $N_{Re}$, the Froude number, $N_{Fr}$, the Power number, $N_P$, or the Weber number, $N_{We}$, may have an influence on the magnitude of $K_La$.

In the analysis of a surface aerated system then, it can be observed that the impeller produces a pumping action, pumping the liquid in the immediate vicinity of the impeller to a higher elevation than the mean liquid level in the tank. This action could be likened to a hydraulic jump. In the vicinity of the impeller, gravity forces must therefore be important. Thus, gravity effects on the liquid system indicate that the Froude number, rather than the Reynolds number, may be the criterion for hydraulic similarity in these systems.

In other types of aeration systems viscous, or surface tension forces may play an important role and thus be the criterion for hydraulic similarity. For the case of a mechanically surface aerated system then, under conditions of similar reactor and impeller geometry, as well as environmental conditions, the original thirteen dimensionless terms in equation (4) are reduced to three as shown in the general equation:

$$fn\left(\frac{HI}{DI}, N_{Fr}, \frac{KLA}{N}\right) = 0 \qquad (5)$$

If one goes one step further and assigns a constant value to the immersion simplex (HI/DI) for both model and prototype, the equation is further simplified. The equation also indicates that $K_La$ as a dependent variable, is a function of the Froude number ($N_{Fr}$) containing the independent variables of impeller speed (N), impeller diameter (DI) and the gravitational constant, g.

$$KLA = C^1 (N_{Fr}) \qquad (6)$$

Equation (6) can be used to describe the oxygen transfer characteristics of each system. For the development of a scale-up prediction model one can explore the suitability of a power relationship such as:

$$KLA = \emptyset N^a DI^b \qquad (7)$$

where:

$N$ = impeller speed ($T^{-1}$)
$DI$ = impeller diameter ($L$)
$\emptyset$, a, b = constants

If one postulates that scale-up of surface agitated (turbine), complete-mixing, unbaffled, square, geometrically similar reactors can be achieved when the overall oxygen transfer coefficient in model and prototype are equal, then a scale-up equation can be developed as follows:

$$(KLA_{20})\ \text{model} = (KLA_{20})\ \text{prototype} \qquad (8)$$

from (7)

$$(N^a DI^b)_{\text{model}} = (N^a DI^b)\ \text{prototype} \qquad (9)$$

which upon simplification reduces to --

$$N_m = N_p \left(\frac{DI_p}{DI_m}\right)^c \qquad (10)$$

where:

$N_m$ = model impeller speed ($T^{-1}$)
$N_p$ = prototype impeller speed ($T^{-1}$)
$DI_p$ = prototype impeller diameter ($L$)
$DI_m$ = model impeller diameter ($L$)
$c = b/a$ = a constant

In order to ascertain whether a scale-up relationship such as equation (10) can be developed and whether conditions for equal process results between different sized but geometrically similar vessels can be generated, a total of 189 experiments using three geometrically similar laboratory-scale model reactors as shown in Figure 1 were conducted. These studies are reported on in detail elsewhere (7). Table 1 indicates the ranges for the experimental conditions imposed for the study.

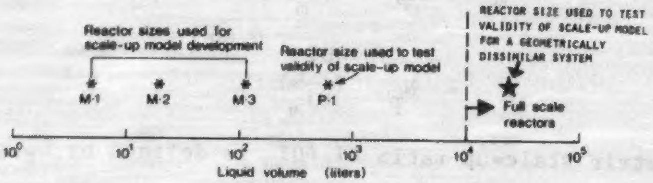
Reactor sizes used for scale-up model development
Reactor size used to test validity of scale-up model
REACTOR SIZE USED TO TEST VALIDITY OF SCALE-UP MODEL FOR A GEOMETRICALLY DISSIMILAR SYSTEM
M·1
M·2
M·3
P·1
Full scale reactors
$10^0$
$10^1$
$10^2$
$10^3$
$10^4$
$10^5$
Liquid volume (liters)

Figure 1. Spectrum of reactor sizes for scale-up model testing.

TABLE 1. RANGES FOR THE EXPERIMENTAL CONDITIONS

| MODEL | NO. OF EXPERIMENTS | ENVIRONMENTAL CONDITIONS |  | HYDRAULIC CONDITION | HYDRODYNAMIC CONDITION |  | NO. OF DIFF. SPEEDS |
|---|---|---|---|---|---|---|---|
|  |  | TEMP $^0C$ | LIQUID MEDIUM | VOLUME LITERS | RPM | $N_{Fr}$ |  |
| M.1 | 106 | 17.70-19.88 | Tapwater | 5.13 | 234-473 | 0.084-0.342 | 57 |
| M.2 | 50 | 17.60-19.33 | Tapwater | 19.7 | 195-331 | 0.091-0.261 | 33 |
| M.3 | 33 | 17.97-19.80 | Tapwater | 113 | 144-202 | 0.090-0.176 | 5 |
| P.1 | 26 | 16.88-21.10 | Tapwater | 607 | 50-132 | 0.019-0.132 | 19 |

## General Scale-Up Model

It had been shown in the analysis of variables influencing mass transfer for this study, that within specific boundary conditions, $K_La$ has a functional dependence on the impeller speed N and the impeller diameter DI. The independent variables N and DI were incorporated in the Froude number ($N_{Fr}$) in equation (6). Further considerations, with the assumption of a power relationship between variables, led to equation (7).

Upon correcting $K_La$ for temperature effects to 20°C, the data from models M.1, M.2 and M.3 were correlated with $K_La_{20}$ at constant immersion depth simplex HI/DI. From 189 observations for 3 models covering a hydrodynamic ($N_{Fr}$) range of 0.084 to 0.342 the following correlation was determined:

$$KLA_{20} = 1.527 \times 10^{-8} N^{2.34} DI^{1.52} \quad (11)$$

The calculated correlation coefficient of 0.9908 accounts for 98% of the correlation by using the two independent variables N and DI.

## Scale-Up Model Development

Based on the postulate of equal mass transfer rates in model and prototype as a prerequisite to the scaling of bio-hydraulic systems, the scale-up equation (10) was developed. Substitution of the exponents for N and DI of equation (11) into the scale-up equation (10) establishes the relationship:

$$N_m = N_p \left(\frac{DI_p}{DI_m}\right)^{1.52/2.34}$$

$$= N_p \left(\frac{DI_p}{DI_m}\right)^{0.65}$$

If the geometric scale-up ratio $DI_p/DI_m$ is defined by $L_r$, then:

$$N_m = N_p L_r^{0.65} \qquad (12)$$

Equation (12) is the scale-up function for the systems examined. $L_r$ may also describe the ratio of any other geometric system variable provided that similar systems are studied. Equation (12) also allows one to predict from prototype data the hydrodynamic condition in a scaled-down model which will give equal $K_La_{20}$ values. Equation (12) was used to predict hydrodynamic conditions which would give equal mass transfer rates between laboratory and pilot-scale model.

Scale-Up Model Testing

In order to test the scale-up model, a larger pilot-scale reactor (P.1) was designed. The reactor liquid volume of approximately 600 liters was selected as a size which is relatively easy and inexpensive to construct. This reactor size was also nearly one order of magnitude larger than the largest laboratory-scale reactor used to develop the prediction model.

The mathematical model for $K_La_{20}$ prediction, equation (11), was tested as shown by example calculation using the laboratory-scale reactor system M.1. For M.1, the impeller diameter ($DI_m$) is 5.40 cm. An impeller speed of 250 rpm was arbitrarily selected. Substituting these values into equation (11), $K_La_{20m}$ is calculated to equal 0.080 $min^{-1}$. The impeller speed ($N_p$) for the model P.1 model which will give a $K_La_{20p}$ value of 0.080 $min^{-1}$ must now be established.

It is known that the pilot-scale reactor impeller diameter $DI_p$ = 26.67 cm. Substituting these data into the equation (12).

$$250 = N_p (4.94)^{0.65} \qquad L_r = \frac{26.67}{5.40} = 4.94$$

$$N_p = 90 \text{ rpm}$$

This means that at impeller speeds of 90 rpm for the pilot-scale reactor (P.1) and 250 rpm in model M.1 the $K_La_{20}$ values should equal 0.080 $min^{-1}$. In order to see if indeed this is true, the calculated prototype speed ($N_p$ = 90 rpm) is substituted into equation (13) which describes the experimentally derived $K_La_{20}$ values for the pilot-scale reactor and then calculating the prototype $K_La_{20}$ value.

The linearized data from the experimental $K_La_{20}$ determinations for the pilot-scale reactor (P.1) may also be described by a power function as:

$$KLA_{20} = 2.57 \ N_{Fr}^{\ 1.244} \qquad (13)$$

The Froude number ($N_{Fr}$) is calculated as:

$$N_{Fr} = \frac{N^2 DI}{g} = \left(\frac{90^2}{60^2}\right)\left(\frac{26.67}{980}\right) = 0.061$$

and substituting into equation (13) yields:

$$KLA_{20} = 2.57 \ (0.061)^{1.244} = 0.080 \text{ min}^{-1}.$$

The actual and predicted $K_La_{20}$ values for P.1 (equations 13 and 11, respectively) are in excellent agreement, thus confirming the validity of the oxygen transfer coefficient prediction equation (11) and the scale-up equation (12). Similar calculations are summarized in Table 2 for geometric ratios of 4.94, 3.18 and 1.75 at three different impeller speeds. Figure 2 illustrates more dramatically the excellent agreement between the in situ and predicted $K_La_{20}$ values for P.1 based on equation (11).

The individual in situ measured $K_La_{20}$ values for P.1 and the predicted values based on equation (11) as developed from M.1, M.2 and M.3 data, are summarized elsewhere (7).

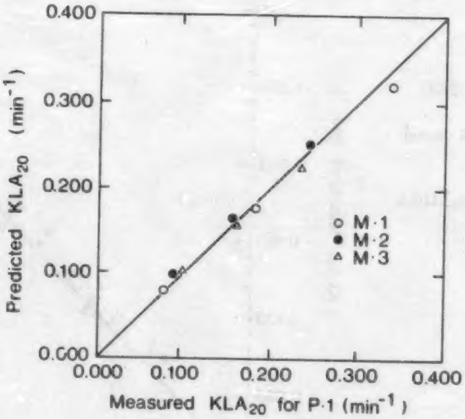

Figure 2. Comparison between in-situ $K_La_{20}$ for P.1 and predicted $K_La_{20}$ from M.1, M.2 and M.3 data

TABLE 2. COMPARISON BETWEEN PREDICTED AND MEASURED $K_La$ VALUES FOR P.1 CALCULATED FROM M.1, M.2 AND M.3 DATA

| MODEL | PARAMETER | | | | | P.1 | | |
|---|---|---|---|---|---|---|---|---|
| | $N_m$ rpm | $DI_m$ cm | $DI_p$ cm | $L_r$ -- | $L_r^{0.65}$ | $N_{pc}$ rpm | $KLA_{20_p}$ (7) $min^{-1}$ | $KLA_{20_m}$ (12) $min^{-1}$ |
| M.1 | 250<br>350<br>450 | 5.40 | 26.67 | 4.94 | 2.795 | 90<br>125<br>161 | 0.080<br>0.177<br>0.322 | 0.080<br>0.182<br>0.340 |
| M.2 | 200<br>250<br>300 | 8.40 | 26.67 | 3.18 | 2.120 | 94<br>118<br>141 | 0.090<br>0.157<br>0.244 | 0.097<br>0.163<br>0.252 |
| M.3 | 140<br>170<br>200 | 15.25 | 26.67 | 1.75 | 1.439 | 97<br>118<br>139 | 0.099<br>0.157<br>0.235 | 0.104<br>0.159<br>0.224 |

Scale-Up Model Refinement

By including the data from P.1 in the mathematical model previously obtained from the M.1, M.2 and M.3 data, a further refinement of the scale-up prediction model from regression analysis resulted in:

$$KLA_{20} = 0.98 \times 10^{-8}\ N^{2.41}\ DI^{1.55} \qquad (14)$$

The multiple regression coefficient increased slightly from 0.9908 for equation (11) to 0.9949 for (14) - explaining 99% of the correlation by using the two independent variables N and DI.

Figure 3 shows the grouped data of 215 observations for all four reactor sizes while comparing predicted to measured values. The high correlation coefficient of 0.9949 is illustrated by the lack of data scatter, even at the extremes.

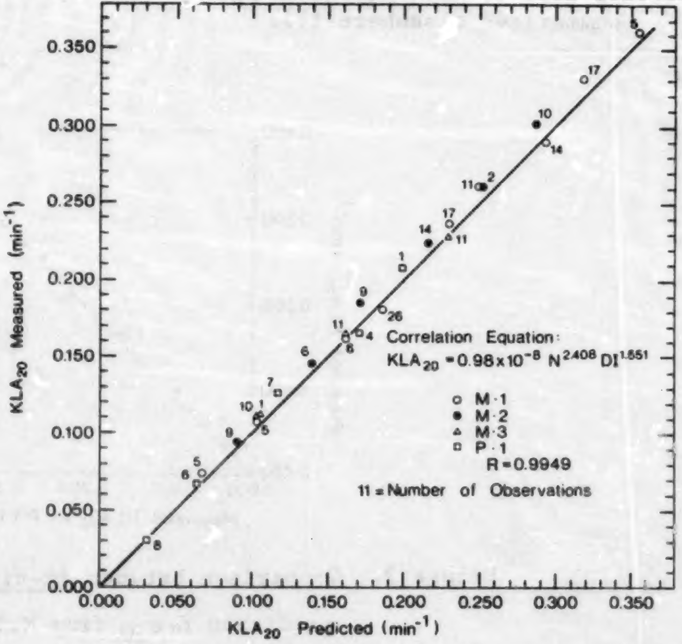

Figure 3. Graphic illustration of correlation equation showing predicted and measured $K_La_{20}$ values for all models

### Error Calculation

During the course of any experimental cbservation, a systematic error is made. In order to assess the magnitude of the possible systematic error that could be included in a calculation of $K_La_{20}$, the partial differential of equation (14) was determined as follows:

$$\frac{\delta KLA_{20}}{KLA_{20}} = 2.41 \ \frac{\delta N}{N} + 1.55 \ \frac{\delta DI}{DI} \qquad (15)$$

The $K_La_{20}$ value for each model at low and high impeller speeds was calculated from equation (14). For all experiments the impeller speed was determined within ±1 rpm while the tolerance for the impeller geometry was ±0.10 cm. Calculations carried out for the maximum and minimum hydrodynamic condition of each of the four models summarized in Figure 4 illustrates that the systematic error for $K_La_{20}$ prediction is directly proportional to the Froude number as:

$$\pm \ \delta KLA = 0.034 \ N_{Fr} \qquad (16)$$

With model M.1 the error in $K_La_{20}$ due to N and DI ranges from ±0.003 $min^{-1}$ at low speed to ±0.013 $min^{-1}$ at high speed. For model P.1, the error at low and high speeds decreased to ±0.001 and ± 0.005 $min^{-1}$, respectively.

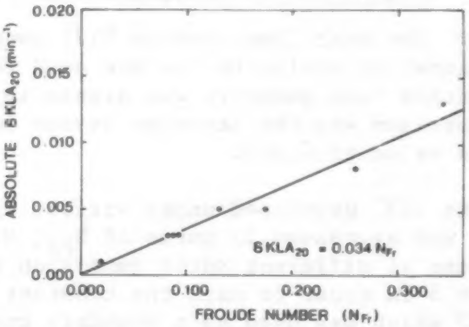

Figure 4. Possible systematic error in $K_La_{20}$ due to N and DI

It would be of considerable importance if equation (12) would also hold for geometrically dissimilar surface aeration systems. As well, the boundary conditions for applying equation (12) would have to be defined.

### Scale-Up for a Geometrically Dissimilar System

The objective of this analysis then, was to examine under what conditions, if any, the scale-up transform

$$N_m = N_p L_r^{0.65} \qquad (12)$$

is applicable to a geometrically dissimilar surface aeration system.

Because the scale-up equation (12) was developed under conditions of a constant aeration device immersion simplex, any comparison with another system must also be made at the same simplex value. It was decided to see if the equation would hold for data (12) concerning oxygenation experiments in tap water to determine the oxygen transfer efficiency of a bladed rotor (cage rotor). These data were developed from studies conducted in a rectangular aeration tank 603 cm in length, 152 cm in width at a liquid depth of 239 cm. The surface aeration device was a bladed rotor, 99 cm in length and 700 mm diameter.

Table 3 summarizes the various geometric simplexes identified in equation (4), sizes them to those used in the development of the scale-up equation (12) and compares them with the geometrically dissimilar system under analysis.

TABLE 3. SUMMARY COMPARISON OF AERATION SYSTEM GEOMETRIC SIMPLEXES

| | Simplex | (7) | (12) |
|---|---|---|---|
| Aeration Tank | Shape | Square | Rectangular |
| | W/DI | 4.0 | 2.17 |
| | W/D | 2.0 | 0.63 |
| Aeration Device | Type | Rushton Turbine | Cage Rotor |
| | WI/DI | 0.20 | 0.07 |
| | LI/DI | 0.25 | 0.15 |
| | S/DI | -- | -- |
| | B | 6 | >>6 |
| Aerator Immersion | HI/DI | 0.202 | 0.202 |

Clearly, not only were the data from studies (12) conducted with a geometrically dissimilar aeration device to the one used for the development of equation (12), but aeration tank geometry was dissimilar as well. The common basis for data comparison was the aeration device immersion simplex (HI/DI) which was set at a value of 0.202.

The reoxygenation data (12) developed under various hydrodynamic conditions were converted and expressed in terms of $N_{Fr}$, HI/DI and $K_La_{20}$. Since the data were obtained at different rotor immersion simplexes, it was plotted as shown in Figure 5 in order to mark the constant aeration device immersion simplex of 0.202 which was used as a boundary condition for the development of the scale-up equation (12). This then was the basis for comparison between the geometrically dissimilar system and the experiments carried out in the earlier study (7).

As indicated on Figure 5 at a constant aeration device immersion simplex of 0.202, $K_La_{20}$ values were selected for various hydrodynamic conditions as defined by $N_{Fr}$. These data are summarized in Table 4 and plotted in Figure 6. The regression equation was calculated as:

$$K_La_{20} = 2.31\ N_{Fr} - 0.056 \qquad (17)$$

where:

$K_La_{20}$ = standard overall oxygen transfer coefficient ($min^{-1}$)
$N_{Fr}$ = Froude number

The Froude number can also be expressed in terms of an aerator peripheral velocity and the liquid volume of the aeration tank. From a practitioners'

TABLE 4. STANDARD OVERALL OXYGEN TRANSFER COEFFICIENTS UNDER VARYING HYDRODYNAMIC CONDITIONS AT HI/DI = 0.202

| N | $N_{Fr}$ | $Kla_{20}$ |
|---|---|---|
| rpm | -- | $min^{-1}$ |
| 59 | 0.069 | 0.118 |
| 73 | 0.105 | 0.189 |
| 88 | 0.153 | 0.265 |
| 100 | 0.198 | 0.410 |
| 120 | 0.285 | 0.610 |

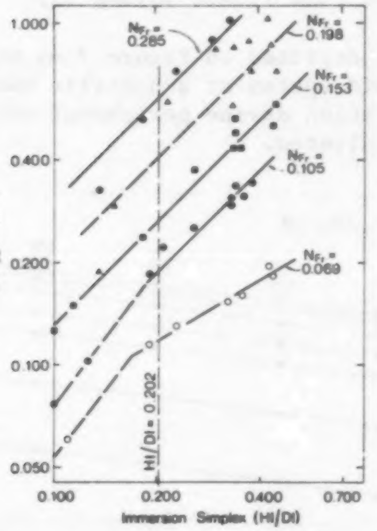

Figure 5. Standard overall oxygen transfer coefficient at various immersion simplexes.

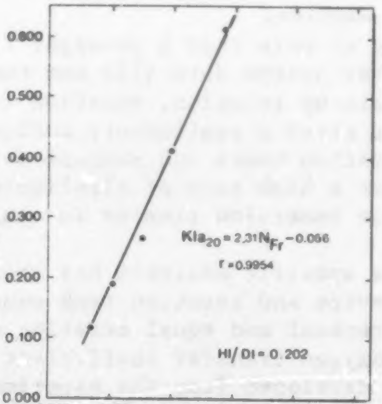

Figure 6. Standard overall oxygen transfer coefficient at various hydrodynamic conditions.

point of view the concept of peripheral velocity rather than Froude number may be more meaningful.

It was determined that the data from all models including the study data (12) can be fitted with the transform of equation (14) as:

$$KLA_{20} = 6 \times 10^{-6} \; V_p^{2.36} \; V^{-0.28} \qquad (18)$$

where:

$V_p$ = the aeration device peripheral velocity ($cm.s^{-1}$)
$V$ = aeration tank liquid volume (l)

These data can be depicted in Figure 7 so that if it is desirable to operate any sized liquid system at a specific overall oxygen transfer value, the surface aeration device peripheral velocity required to attain this $K_La_{20}$ value is indicated.

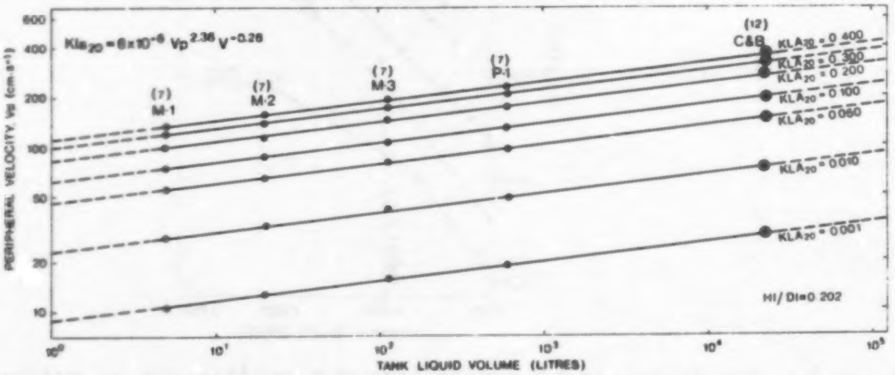

Figure 7. Surface aeration device peripheral velocity and standard overall oxygen transfer coefficient for various liquid volumes at constant immersion simplex.

It is interesting to note that a straight line correlation between the geometrically dissimilar system data (12) and the data used in the development of the scale-up relation, equation (12) exists. This is highly significant in that it gives a preliminary indication that geometric similarity for the aeration tanks and surface aeration devices may not be required, provided that a high rate of air-liquid interface renewal occurs and the aeration device immersion simplex is constant between all systems.

Essentially, this specific analysis has shown that, for a geometrically dissimilar aeration device and aeration tank under conditions of intensive air-liquid interface renewal and equal aeration device immersion simplex (HI/DI), the overall oxygen transfer coefficient ($K_La_{20}$) could be predicted from equation (14) as developed from the experimental work (7). The inference that can be drawn from this is: while system geometry may not be critical, the degree of turbulence generation and aeration device immersion simplex is critical. Cage rotors have these characteristics.

### Recent Surface-Aeration Scale-Up Study

Extensive investigations and review concerning surface-aerator scale-up were recently reported on (10). With a scale factor of 10 between two aerators it was observed that the deviation in oxygen transfer results between small and large scale systems was within ±20%. This was considered to be acceptable. Encouraged by these results, it would appear then that a similar approach can be taken in scaling other aeration systems such as diffused, turbine (sparger), static and rotors when equal environmental conditions between model and prototype exist.

## $K_La$ AND SYSTEM VARIABLES FOR OTHER AERATION SYSTEMS

### $K_La$ and Diffused Aeration System Variables

For a diffused aeration system the oxygen transfer capability may be expressed as a function of gas flow (G), liquid depth above the diffusers (H) and the width of the aeration tank (W) as:

$$K_La = fn\ (G, H, W) \qquad (19)$$

### $K_La$ and Turbine (Sparger) Aeration System Variables

One would suspect that for a turbine aeration system the overall oxygen transfer coefficient is a function of the combined variables for diffused aeration and surface aeration. This indeed is true and can be expressed as:

$$K_La = fn\ (G, H, N, DI) \qquad (20)$$

### $K_La$ and Static Aeration System Variables

A static aeration system consists of submerged vertical cylindrical tubes which may contain a helix to lengthen the bubble path and hence contact time. Compressed air is fed through a sparger at the cylinder bottom. Variables affecting $K_La$ are gas flow (G), sparger design (K), cylinder length (L) and diameter (D) and liquid depth (H). This may be expressed in a functional relationship as:

$$K_La = fn\ (G, L, D, H) \qquad (21)$$

### $K_La$ and Cage Rotor Aeration System Variables

A rotor aerator could be likened to a turbine aerator on its side and thus have variables influencing the overall oxygen transfer coefficient $K_La$ which are identical:

$$K_La = fn\ (N, DI, HI) \qquad (22)$$

This appears to be substantiated by the analysis of data (13) between two geometrically dissimilar aeration systems summarized earlier.

Studies have shown (14) that $\alpha$ is also a machine variable, being higher for surface than for diffused aeration systems. It would be appropriate to highlight some of the studies conducted by others concerning scale-up.

Diffused Aeration System Scale-Up

Using dimensional analysis as a starting point a general relationship for the prediction of the overall oxygen transfer coefficient, $K_La$ was developed (5):

$$\frac{K_La}{Q_a} = 9.35\left[\frac{\rho Q_a(Nd_o)^2}{\mu_L}\right]^{0.046}\left[\frac{Q_a{}^2(Nd_o)}{g}\right]^{-0.126}\left[\frac{d_d}{T_d}\right]^{0.69}\left[\frac{T_w}{T_d}\right]^{-0.446} \qquad (23)$$

where:

$Q_a$ = airflow/unit volume $T^{-1}$
$\rho$ = density of liquid $FL^{-4}T^2$
$\mu$ = viscosity of liquid $FL^{-6}T^3$
$N$ = number of air bubbles --
$d_o$ = diffuser orifice diameter $L$
$g$ = gravitational constant $LT^{-2}$
$d_d$ = diffuser depth $L$
$T_d$ = aeration tank depth $L$
$T_w$ = aeration tank width $L$

The coefficients for the correlation were developed from reaeration experiments conducted in 3 geometrically similar but different sized aeration tanks. The equation describes his data with a mean error of less than 10 percent.

The same researcher (5) also developed a general relationship relating key parameters directly to $K_La$.

$$K_La = 28.60(G)^{0.82}\left[\frac{1}{V}\right]^{1.06}(d_d)^{0.724} \qquad (24)$$

He found that this equation predicted $K_La$ for his data with a mean error of less than 8%.

Turbine (Sparger) System Scale-Up

Another researcher (9) investigated parameters influencing $K_La$ for a turbine (sparger) system. While his correlations described his data, successful scale-up remained elusive.

Static Aeration System Scale-Up

The problem of bench and full scale tests for alpha and beta coefficient variability determination was recently addressed (14). As shown in Figure 8 which compares bench and full scale $K_La$ values with corresponding airflow per unit volume, approximately 2.5 to 3 times more air is required for the bench scale unit than for the prototype unit in order to produce equal $K_La$ values. The primary reason for this is the difference in liquid depth between the two systems. i.e. 90 cm vs 305 cm. This would suggest that distortion of the model geometry to allow for equal bubble hold-up time may be in order. A scale-up function was not developed in this work (14).

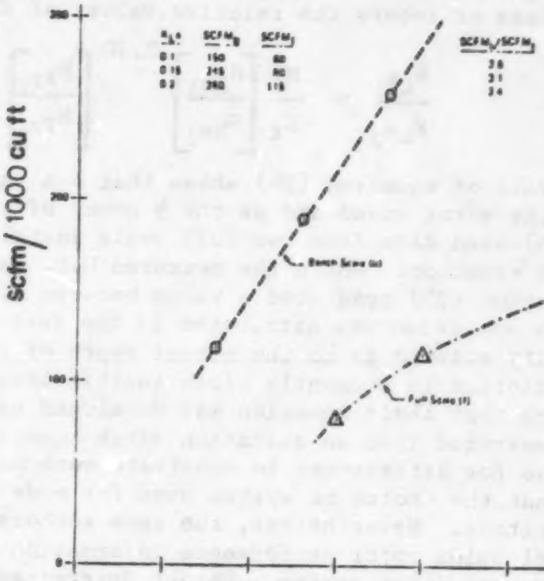

Figure 8. Comparison between full-scale and bench scale $K_La$ at unit air flows.

## Cage Rotor Aeration System Scale-Up

An interesting study (6) reports on the successful development of a predictive equation for oxygen transfer using cage rotors. These researchers started with bench top size models and validated their predictive equation using data from existing full-scale installations at Aberdeen, Scotland and Ames, Iowa. The predictive equation derived with the assistance of dimensional analysis was:

$$\frac{K_La}{N} = 7.42\times10^{-7}\left[\frac{HI}{DI}\right]^{0.86}\left[\frac{P_W}{DI}\right]^{-0.18}\left[\frac{H}{DI}\right]^{-0.28}\left[N_{Re}\right]^{0.70}\left[N_{Fr}\right]^{-0.19} \qquad (25)$$

where:

| | | | |
|---|---|---|---|
| $K_La$ | = | overall oxygen transfer coefficient | $T^{-1}$ |
| N | = | rotor speed | $T^{-1}$ |
| HI | = | depth of rotor immersion | L |
| DI | = | rotor diameter | L |
| $P_W$ | = | rotor blade width | L |
| H | = | liquid depth | L |
| $N_{Re}$ | = | dimensionless Reynolds Number | - |
| $N_{Fr}$ | = | dimensionless Froude Number | - |

Equation (25) was developed and coefficients evaluated using a single 1/12th scale model operated under a great variety of operating conditions. Correlation coefficients for this data varied from 0.983 to 0.996 for each of the components. The equation shown above can be put into a more general form and used for scale-up purposes to estimate oxygenation capacities for rotors in general even though system geometry differs. For a geometrically similar class of rotors the relative values of $K_La$ are:

$$\frac{K_La_1}{K_La_2} = \frac{N_1}{N_2}\left[\frac{N_{Re_1}}{N_{Re_2}}\right]^{0.70}\left[\frac{N_{Fr_1}}{N_{Fr_2}}\right]^{-0.19} \qquad (26)$$

Analysis of equation (26) shows that $K_La$ varies directly as the 4/3 power of the rotor speed and as the ½ power of the rotor diameter. The authors (6) used data from two full scale installations to validate their predictive equation. While the measured O.C. was 2.15 lbs oxygen per day their equation (25) predicted a value between 2.26 and 2.37 lbs oxygen per day. This variation was attributed to the fact that some degree of uncertainity existed as to the actual depth of rotor immersion. Nevertheless their prediction is uncannily close (within less than 11%), especially when considering that their equation was developed using pure water and compared to data generated from an oxidation ditch treating swine wastes. No corrections for differences in substrate were made. This leads one to suspect that the choice of system used for model validation might have been most fortuitous. Nevertheless, the same authors (6) went on to validate their model using rotor performance information generated (15) for a full-scale oxidation ditch system. Knight determined that after one hour aeration of the initially deoxygenated ditch water, the dissolved oxygen concentration had increased to 6.33 $mgL^{-1}$. Application of equation (25) estimated a dissolved oxygen concentration of 7.08 $mgL^{-1}$. This again represents agreement to within less than 12%.

## Rotating Biological Contactor Aeration Scale-Up

A very preliminary stab at examining oxygen transfer for RBC can be made using limited data (16). Again one can consider that $K_La$ is some function of the contactors' rotational speed, N, and its diameter, DI. As well, disc immersion depth, HI, is important. A function describing oxygen transfer in an RBC system could then be written as:

$$K_La = \text{fn}\ (N, DI, HI) \qquad (27)$$

It can be postulated that a scale-up function at equal immersion simplex assumes the form:

$$N_m = N_p \left(\frac{DI_p}{DI_m}\right)^c \qquad (28)$$

and that:

$$K_La = \phi N_{Fr}^{\ n} \qquad (29)$$

**Table 5 summarizes the oxygen transfer data calculated to produce Figure 9. The data are most interesting in that they illustrate that for this particular immersion depth simplex (constant between models) a limiting overall oxygen transfer coefficient is attained. It would appear from this albeit sparse data that the maximum value of the overall oxygen transfer coefficient is around 0.050 min$^{-1}$.**

**TABLE 5. RBC OXYGEN TRANSFER DATA AT DIFFERENT HYDRODYNAMIC CONDITIONS**

| DI<br>cm | N<br>rpm | $V_p$<br>cm s$^{-1}$ | $N_{Fr}$<br>- | $K_La$<br>min$^{-1}$ |
|---|---|---|---|---|
| 15.25 | 40 | 33 | 0.0074 | 0.049 |
| | 20 | 16.5 | 0.0018 | 0.036 |
| | 10 | 8.25 | 0.0005 | 0.021 |
| 30.5 | 20 | 33 | 0.0037 | 0.042 |
| | 10 | 16.5 | 0.0009 | 0.031 |
| | 5 | 8.25 | 0.0002 | 0.011 |
| 61 | 10 | 33 | 0.0018 | 0.028 |
| | 5 | 16.5 | 0.0005 | 0.015 |
| | 2.5 | 8.25 | 0.0001 | 0.008 |

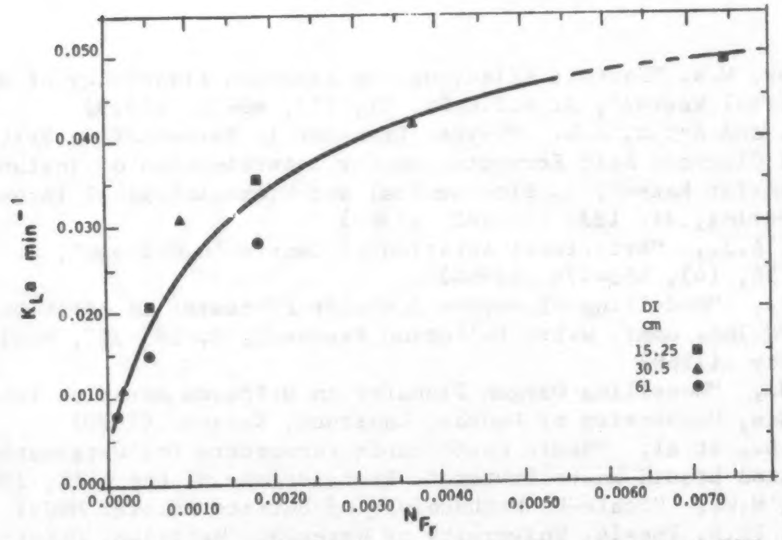

**Figure 9. $K_La$ as a function of $N_{Fr}$ for different sized RBC systems**

**Information and correlations of this type would greatly assist in the generation of process design and scaling information. Admittedly only the surface has been scratched.**

## SUMMARY

The information presented clearly indicates that progress in the area of process scaling has been made. Specifically, our ability to predict full scale oxygen transfer capabilities of various aeration devices and systems from small scale systems has increased. But it is equally apparent that the amount of data scatter in investigations concerning system scale-up, lack of suitable field data for model validation is still formidable. Differences between various formulations are equally extreme.

It is evident that we still have a long way to go before oxygen transfer data generated at bench-scale can be used with great confidence for prototype aeration system scale-up. Equally, the transfer functions for scale-down remain to be developed and validated.

There is little doubt that in order to minimize scale-effects between laboratory and prototype systems every attempt at attaining the goal of environmental, geometric, kinematic and dynamic similitude must be made. Successful scaling correlates directly with the degree to which we attain this goal.

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# AERATION SYSTEM DESIGN PROTOCOLS: A NORTH AMERICAN PERSPECTIVE

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## ABSTRACT

Aeration system sizing for suspended growth systems requires careful evaluation of organic and hydraulic loading parameters. Experience based and rationally developed model alternatives for estimating oxygen demand are presented. Physical requirements and conditions, including mixing, tank geometry and aeration system type and layout are covered. The sensitivity of the design to various parameters is also investigated. The results of a small survey on aeration system design approaches is included.

## INTRODUCTION

All biological treatment processes make use of oxygen during the metabolic steps involved in the conversion of the organic matter to cellular mass components and energy. The design of biological treatment systems can be accomplished making use of varying amounts of knowledge of the system biochemistry. The objective of waste treatment research is to achieve a better understanding of the biochemical processes involved and, using this information, develop rational design relationships for sizing of process reactors and components. A variety of such approaches have been followed in the development of aerobic suspended growth design models. The application of such models requires the user to determine a number of parameters for the design of wastewater treatment systems.

This discussion will focus on the approaches used in North America for ultimately identifying the oxygen requirement in aerobic suspended growth systems. Topics to be covered include estimating the organic loads and its variations; estimating oxygen requirements and its variations; the effects of tank geometry, aeration system layout, and mixing; the sensitivity of the design to various parameters; and the results of a survey on current practice.

## SYSTEM PERFORMANCE MODELLING

The development of models for suspended growth system performance prediction have taken two basic forms:

- experience based rules for design and operation, and
- rationally developed models for design and operation.

For each of these approaches, the following basic informational components are required:

- organic loading and its variation with time,
- hydraulic loading and its variation with time,
- nutrient complement and its variation with time, and
- definition of the physical environment.

Each of these components will vary with the hour of the day, the day of the week, the week or month of the year, and from year to year. For municipal systems the pattern of variation is dependent on the social and industrial infrastructure of the community. For design purposes the values and variations at the time of design and, more importantly, the future projections of the values to the design life of the facility are required.

## Organic Loading

The organic loading to a treatment plant can be evaluated using one of the following parameters:

- biochemical oxygen demand (BOD)
- chemical oxygen demand (COD)
- total organic carbon (TOC), and
- total oxygen demand (TOD)

Each of these measurements can be related to the substrate requirements of the microorganisms in the aeration system. However, the $BOD_5$ value of the wastewater most closely relates to the actual oxygen requirements for the metabolic processes involved during the conversion of the useable organics to cell components. Therefore, the design parameter used for expressing the organic load to a wastewater suspended growth treatment process is the BOD. In some cases, only the soluble fraction is considered, however, the realistic approach is to include the total primary effluent $BOD_5$ as the design value. One method of evaluating existing operating conditions is to make use of frequency distributions to describe organics and suspended solids concentration to the treatment unit. Data points used for such distributions should assist in identifying loading rates for design. The optimum approach is to use BOD data average the maximum significant design retention time of the aeration tank. Figure 1 presents a typical frequency distribution for $BOD_5$, COD and total suspended solids for raw wastewater. The daily variation is represented in Figure 2.

The reduction in the raw wastewater by the primary sedimentation process normally varies from 20 to 35 percent for $BOD_5$ and 50 to 65 percent for suspended solids.

If nitrification is allowed to occur in the aeration tank then this oxygen requirement must be estimated and allowed for in the design of the aeration equipment.

## Hydraulic Loading

The hydraulic flow rate through an activated sludge unit is equal in importance to the organic loading. Three flow values are important to design: average, design, and peak. Here the design flow rate will be assumed. This value is that of some maximum significant period such as 4, 6 or 8 hours, depending on the actual conditions. Figure 3 illustrates the type of flow pattern common to municipal systems. Again this must be estimated for the design life of the system and/or design stages.

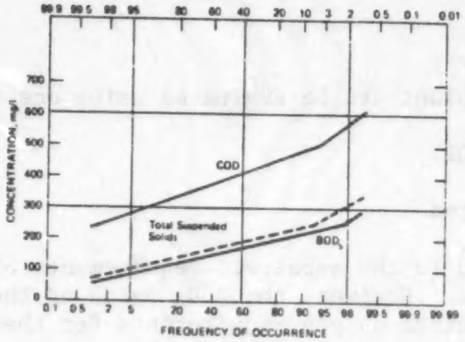


**Figure 1:** Typical frequency distribution for domestic wastewater (after ref. 2)

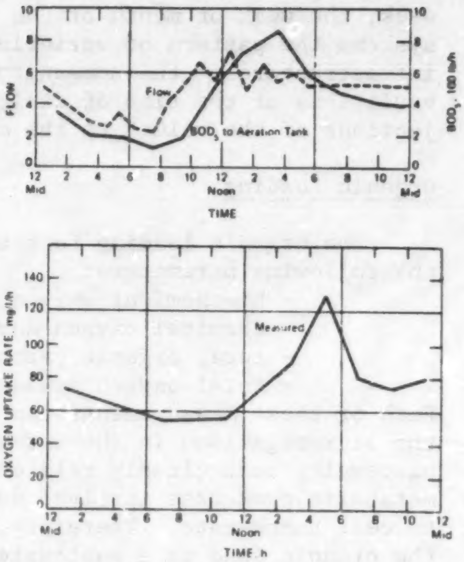


**Figure 2:** Typical daily variation in wastewater flow and $BOD_5$ (after ref. 2)

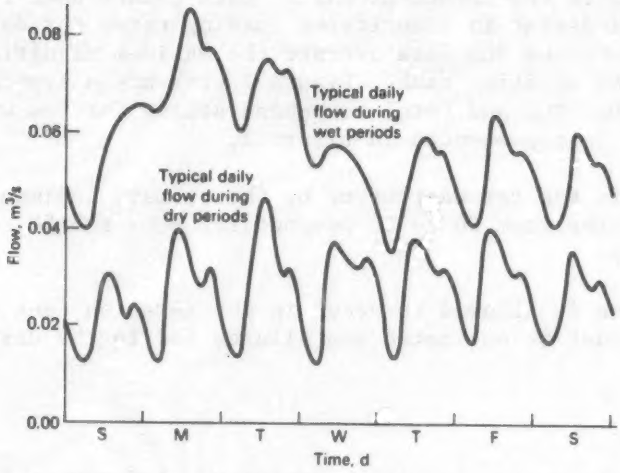


**Figure 3:** Typical municipal wastewater flow pattern (after ref.8)

Nutrient Complement

A large number of nutrients are required for the optimum growth of a biological population. Municipal wastewater is usually adequately supplied with nutrients. However, industrial and mixed industrial and domestic wastewaters may be deficient in some nutrient. This should be investigated prior to design.

## Physical Environment

Although many factors impact on the physical environment of a biological system, temperature is among the most important. Temperature estimation must be incorporated into the information estimating process since it effects the rate of biological activity and the rate of oxygen transfer. Other physical components such as tank geometry and mixing will be discussed later.

## EXPERIENCED BASED ESTIMATES OF OXYGEN DEMAND

A variety of experienced based design relationships have been used in North America. These have varied from 1.0 kg $O_2$ per kg carbonaceous $BOD_5$ removed to relationships relating air to be supplied to $BOD_5$ applied to more elaborate functions of $BOD_5$ and solids retention time. Air application rates for the more frequently used activated sludge processes are shown in Table 1.

A more comprehensive approach was proposed in 1954 by Eckenfelder and O'Connor which related the oxygen requirement to the $BOD_5$ removed and the mass of mixed liquor suspended solids (MLSS) in the aeration tank (3). In SI units the equation takes the form

$$\frac{\text{kg } O_2 \text{ supplied}}{\text{day}} = A'\left(\frac{\text{kg } BOD_5 \text{ removed}}{\text{d}}\right) + B' \text{ (kg MLSS in aeration tank)}$$

where

$A'$ = $O_2$ required for $BOD_5$ synthesis, mg $O_2$/mg $BOD_5$ removed
and
$B'$ = $O_2$ required for endogenous respiration, mg $O_2$/(mg MLSS·d)

Several other researchers also developed values for A' and B' following this relationship. Converted values are shown in Figure 4.

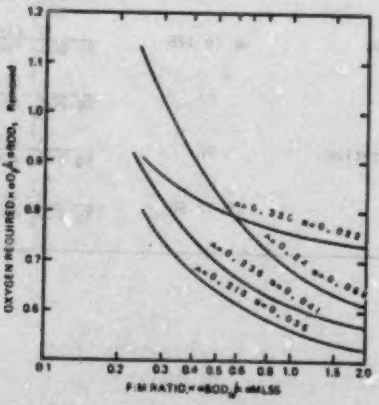


Figure 4: Relationship between F:M ratio, oxygen requirements and different predictive relationships (after ref. 2)

Stensel and Shell related oxygen requirements to $BOD_5$ removed and solids retention time (SRT) (4). Their relationship was of the form:

$$r_0 = \left[\frac{\Delta BOD_5 \cdot Q}{(1 + b)\ SRT}\right] \left[A + bA\ (SRT) + aB\ (SRT)\right]$$

where

$r_0$ = oxygen uptake rate, mg/(L·d)
a = cell yield coefficient, mg cell/mg $BOD_5$ (range of value = 0.5 to 0.7)
b = endogenous decay coefficient, 1/d (range of value = 0.04 to 0.1)

The effect of temperature on the oxygen requirement is shown in Figure 5 for a domestic wastewater.

Table 1
Air Application Rates for Various Activated Sludge Options

| Activated Sludge Alternative | Gas Applied | Units | Reference |
|---|---|---|---|
| Convention | 1.0 | $\frac{Kg\ O_2}{Kg\ BOD_5}$ | 2 |
| | 3.7 to 15 | $\frac{m^3\ air}{m^3\ wastewater}$ | 2 |
| | 30 to 55 | $\frac{m^3\ air}{Kg\ BOD_5\ applied}$ | 2 |
| | 94 | $\frac{m^3\ air}{Kg\ BOD_5\ applied}$ | 1 |
| Extended Aeration | up to 125 | $\frac{m^3\ air}{Kg\ BOD_5\ removed}$ | |
| Step Aeration | 90 | $\frac{m^3\ air}{Kg\ BOD_5\ applied}$ | 2 |
| Contact Stabilization | 90 | $\frac{m^3\ air}{Kg\ BOD_5\ applied}$ | 2 |
| High Rate | 25 to 90 | $\frac{m^3\ air}{Kg\ BOD_5\ applied}$ | 2 |

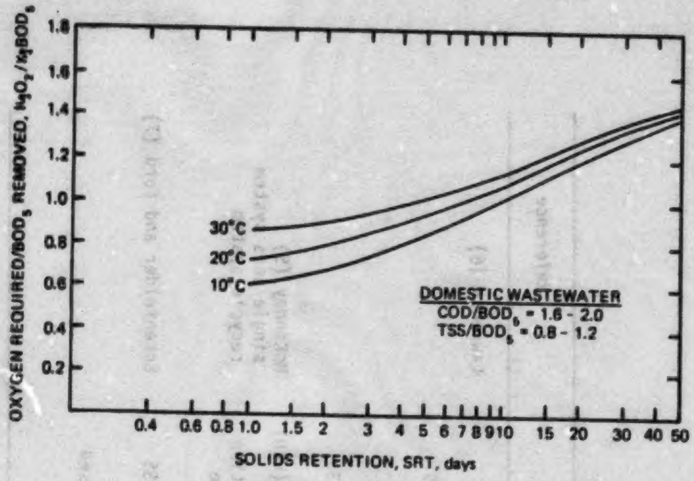


Figure 5: Relationship between solids retention time, temperature and oxygen required per unit $BOD_5$ removed (after ref. 2)

RATIONAL ESTIMATES OF OXYGEN DEMAND

A number of performance and design models have been presented during the last 20 years (5,6,7). The fundamental approach followed by most of the models is the same. This discussion will be limited to the model components related to oxygen uptake rate as developed by McKinney (5), Lawrence and McCarty (6) and Eckenfelder (7). The fundamental difference between these models lies in the relationship between substrate concentration and the growth rate of microbial mass. Discussions of these differences are presented elsewhere (8,9).

The objective of these rational models is to provide a reference base for sizing aeration tanks and equipment, and predicting sludge production and effluent substrate concentration. Table 2 presents the equation for oxygen uptake rate proposed by the three models included here.

The questionnaire responses indicated a variable amount of interest in these models as shown in Table 3.

Table 2. Oxygen Demand Models

| Model | Terms | Reference |
|---|---|---|
| $\Delta O_2 = Q\,[1 - 1.42\,Y_T(S_o-S_e)] + 1.42k_dXV_a$ | $\Delta O_2$ = oxygen demand, g/d<br>$Q$ = flow rate, $m^3/d$<br>$Y_T$ = solids yield coef. mg/mg<br>$S_o$ = infl. substrate, mg/L<br>$S_e$ = effl. substrate, mg/L<br>$K_d$ = end, resp. factor, 1/h<br>$X$ = MLVSS, mg/L<br>$V_a$ = aeration tank volume, $m^3$ | Lawrence (6) |
| $r_r = \frac{1.5\,(S_o-S_e) - 1.42\,(M_a+N_e)}{t}$<br>$r_r = \frac{1.5\,(S_o-S_e)}{t} - \frac{1.42\,(M_a+M_e)}{\theta_c}$ | $r_r$ = oxygen uptake rate, mg/(L•h)<br>$M_a$ = active mass in aeration<br>$M_e$ = endogenous mass in aeration<br>$t$ = hydraulic detention time<br>$\theta_c$ = solids retention time | McKinney (5)<br>single pass system<br>recycle system |
| $r_r = 1.42\,b_oX_d + a'S_r$ | $b_o$ = fraction of degradable SS oxidized per day<br>$X_d$ = degradable MLVSS<br>$a'$ = fraction of substrate used for oxidation<br>$S_r$ = substrate removed | Eckenfelder and Ford (7) |

## Table 3
## Summary of Aeration System Design Survey
## (15 completed responses)

| | |
|---|---|
| **Aeration Systems Used** | |
| Diffused Air | |
| fine bubble | 15 |
| coarse bubble | 66 |
| Mechanical | |
| surface turbine | 56 |
| brush | 15 |
| submerged turbine | 10 (+3 combined) |
| Other | |
| lagoon tubing | 2 |
| static | 11 |
| jet | 3 |
| $O_2$ | 3 |
| | $\Sigma$=184 |

**Design Data Generation (number using indicated approach)**

| Parameter | Literature | Field Measurement | Calculation |
|---|---|---|---|
| Hydraulic | 5 | 12 | 9 |
| BOD | 7 | 11 | 9 |
| Suspended solids | 7 | 12 | 9 |
| Surface active agents | 4 | 7 | 0 |
| Load variation | 9 | 13 | 5 |

**Oxygen Uptake Rate Estimation**

| Method | Number |
|---|---|
| Air Supply/unit BOD | 6 |
| Rational Models | |
| Lawrence and McCarty | 4 |
| McKinney | 4 |
| Eckenfelder | 4 |
| Own | 1 |
| Codes | 4 |

**Aeration Basin Geometry**
Ratio = Length:Width:Depth

Plug flow
6 to 20 : 2 to 3 : 1, sidewater depth 3-5.5 m

| | |
|---|---|
| Aerator pattern | |
| Roll | 5 |
| Grid | 3 |

Complete mix
L:W = 1 to 3 : 1
side water depth 3 - 4.6 m
aerator - mechanical
Circular tanks only noted for package plants

| | | |
|---|---|---|
| **Mixing** | | |
| Minimum liquid velocity | | |
| 0.15 m/s | | 5 |
| basin turnovers | 12-15 times/h | 1 |
| dye testing | | 1 |
| power/unit volume | | 1 |
| air rate/unit volume | | 2 |
| scour calculation | | 2 |
| velocity meter | | 2 |

| | |
|---|---|
| **Process Control** | |
| Use of D.O. probes | |
| Number | |
| 2/tank | 3 |
| 1/tank | 5 |
| Location | |
| point of minimum D.O. | 1 |
| 1/3 point from inlet | 2 |
| midway from aerator to wall | 1 |
| adjustable | 1 |
| Other Methods | |
| portable D.O. probes | 3 |
| field $O_2$ uptake | 1 |
| Winkler test | 1 |
| Feedback Loops | |
| record D.O. only | 2 |
| record D.O. manual control | 3 |
| record D.O. feedback loop | 8 |

## AERATION EQUIPMENT PERFORMANCE

Establishment of the oxygen uptake rate is one of the first steps in the design process. Another early step is establishment of the oxygen delivery capability of alternative types of aeration equipment. The amount of oxygen transferred by an aeration device must be expressed in a form which allows comparison with other units. Oxygen transfer tests are normally conducted with clean water and the results adjusted to standards conditions of 20°C and one atmosphere pressure (2, 11). This type of testing, when conducted in tanks of the same shape and volume, yields information which can be used to compare aerator oxygen transfer efficiency. Performance values are expressed in units of $KgO_2$ transferred per kilowatt-hour (wire to water basis).

Using the clean water non-steady state testing procedure (11) and solving the general equation

$$\frac{dc}{dt} = k_La\,(C_s - C_L)$$

where

$$\frac{dc}{dt} = \frac{\text{change in } O_2 \text{ concentration}}{\text{time}}$$

$K_La$ = overall oxygen transfer coefficient, per hour

$C_s$ = saturation conc. of $O_2$ in water under test conditions, mg/L

$C_L$ = concentration of $O_2$ in water at corresponding time, mg/L

for $K_La$, yields

$$k_La_T = \frac{\ln \dfrac{C_s - C_1}{C_s - C_2}}{t_2 - t_1}$$

which can be used to determine $k_La$. Temperature adjustment is made using

$$k_La_T = k_La_{20^\circ C}\ \Theta^{T-20}$$

where

$T$ = test temperature, °C

$\Theta$ = temperature correction coefficient, 1.024

When the information is for comparison of two units tested in the same tank then $k_La$ or a calculated oxygen transfer rate value in kg per hour can be used. When the information from tests in different size and shape tanks must be used then comparison becomes much more difficult. It has been suggested (12) that the term ($K_La$ · tank volume) can be used. This leaves only the interface area and the mixing factors to be evaluated.

When the results are to be translated into the wastewater system it is necessary to use α and β to adjust for the effects of organics on the wastewater on $k_La$ and of salts, organics and solids on $C_s$, respectively. The general equation then becomes

$$\frac{dc}{dt} = \alpha k_La (\beta C_s - C_L)$$

This equation can be changed in a variety of ways to account for temperature and units adjustments (2).

For development of a design it is necessary to predict the value of α, the correction factor for obtaining a wastewater $k_La$ value, and β, the correction factor for obtaining a wastewater $C_s$ value. Determination and evaluation of these factors has been discussed in detail elsewhere (13). It is known that the values vary from one type of wastewater to the next, from one degree of treatment to another and from one type of aeration equipment to another. For municipal wastewater values for α and β can be selected from the literature with careful consideration of the conditions during which they were measured. Only a limited amount of information has been published regarding industrial waste values of these factors. Normally, a conservative (or low) value should be selected. Due to the problems with performance prediction caused by different aeration basin geometry and with α and β, it is recommended that the aeration equipment specifications provide for either pre-start-up testing or testing within the first year of start-up. Furthermore, penalties for failure to meet oxygen transfer requirements should be specified. Specifications and testing procedures are discussed elsewhere (14, 15, 16, 17).

## PHYSICAL COMPONENTS

The physical components of an aeration system have a major impact on the overall performance of the aeration device used as well as the performance of the biological process. The tank geometry and aeration system layout within the tank control the flow regime and mixing within the reactor.

### Tank Geometry

The extremes of tank geometry are from long, narrow, nearly tubular tanks for a plug flow regime to circular, complete mixing flow regime tanks. These extremes allow easy modelling of mixing and prediction of aeration system design requirements. With modifications to tank configuration, these two extremes are used almost exclusively in aeration tank design.

The mixing patterns in complete mixing reactors is controlled by the aeration pattern. Perfect mixing is difficult and operationally expensive to achieve. Surface or submerged turbine units are frequently used. With mechanical aerators point feed and discharge can be used. With diffused air systems a diffused feed and discharge design is used. Tank configurations vary from circular to square to rectangular about 3:1 length to width ratios.

Plug flow reactors range from 5 to 10 m in width, 5 m in depth and up to 125 m in length. Diffused air systems are frequently used along with point input and outlet. The survey results generally conform with these values.

## Aeration System Layout

The arrangement of the aeration equipment has a major effect on achieving the desired mixing pattern. Figure 6 shows a few alternative arrangements. The wide variety of equipment, tank configurations and arrangements makes comparison of systems difficult.

## Mixing

Mixing has two basic purposes, maintaining the solids in suspension and assuring that the oxygen is well distributed throughout the tank. For some types of aeration devices it is possible to provide adequate oxygen to all parts of the tank without providing adequate suspension of solids and without complete mixing. Complete mixing assumes that the influent is spread throughout the aeration tank in a very short time relative to the retention time. Some researchers have proposed using turnover time (18) or pumping capacity (19) as a measure of the mixing achieved in a reactor. Both values fall short of good description of the operation without specification of baffling and other physical features of the system. Power per unit volume is also a poor indicator of mixing (2).

Mixing is best indicated by tracer studies. Another good indicator in complete mix reactors is the oxygen uptake rate. In an ideally mixed tank the value would be the same at all points. Suspended solids measurements does not provide a good indication of mixing except in extremely poorly mixed conditions.

## STRATEGIES FOR OPERATION AND CONTROL

The design of the aeration system for an activated sludge plant will be significantly influenced by the method of operation selected. The most consistant performance of a biological system can be achieved with uniform organic, hydraulic and recycle sludge loading. This can be approached with flow equalization and the use of parallel reactors with hydraulic loading control.

The aeration system lends itself to continuous or online control by use of a dissolved oxygen monitor linked by a data processor to the drive motors or control valves. To achieve efficient operation variable speed motors or adjustable blade turbines should be used.

The information survey indicate the major of those responding felt feedback loops was desireable. Some respondants felt operator checking and adjusting was more practical.

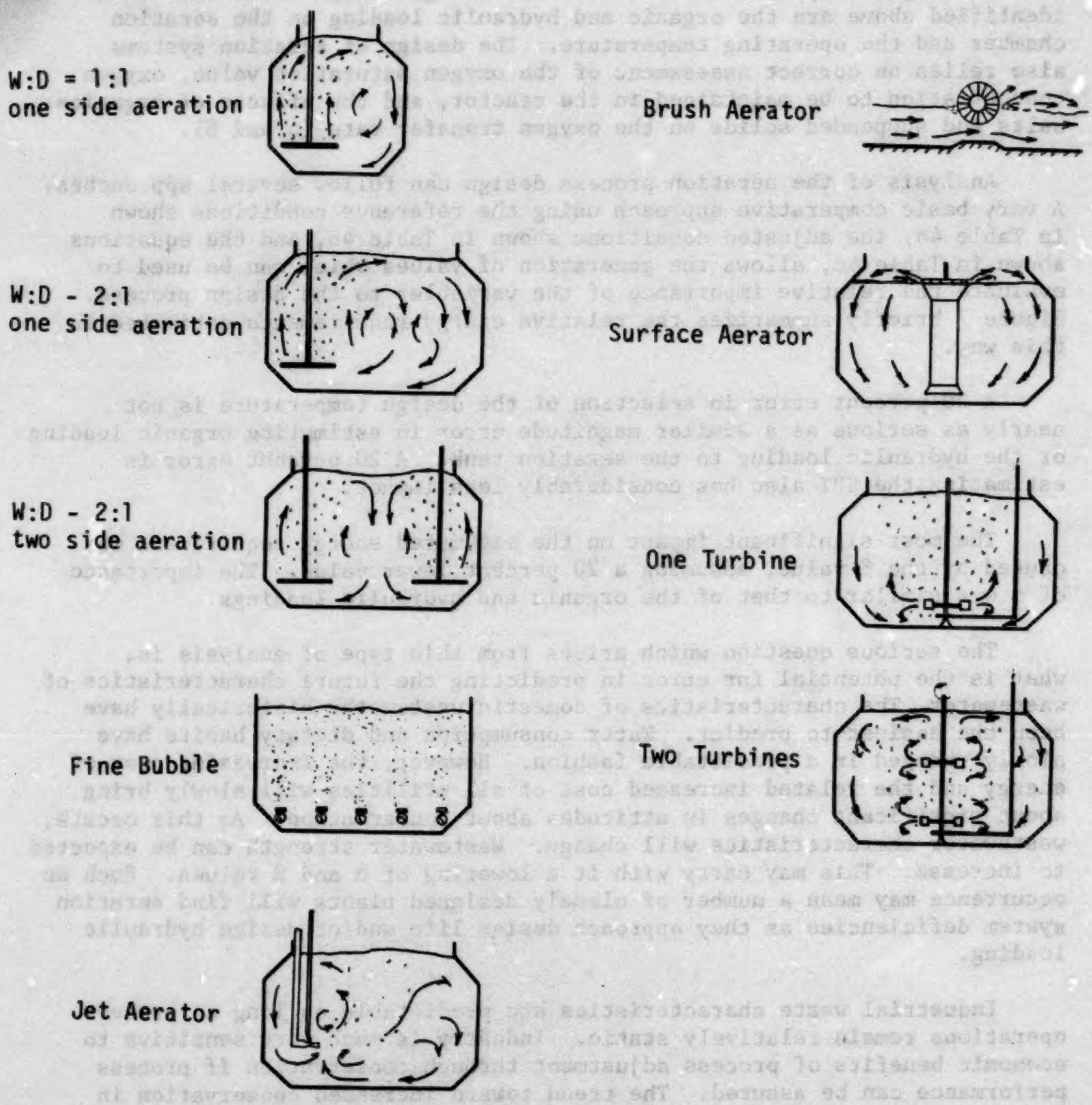

W:D = 1:1
one side aeration
Brush Aerator
W:D - 2:1
one side aeration
Surface Aerator
W:D - 2:1
two side aeration
One Turbine
Fine Bubble
Two Turbines
Jet Aerator

Figure 6: Aeration tank and equipment layouts

## SENSITIVITY OF DESIGN PARAMETERS

The major components of an activated sludge design which have been identified above are the organic and hydraulic loading on the aeration chamber and the operating temperature. The design of aeration systems also relies on correct assessment of the oxygen saturation value, oxygen concentration to be maintained in the reactor, and the effects of organics, salts and suspended solids on the oxygen transfer rate ($\alpha$ and $\beta$).

Analysis of the aeration process design can follow several approaches. A very basic comparative approach using the reference conditions shown in Table 4a, the adjusted conditions shown in Table 4b, and the equations shown in Table 4c, allows the generation of values which can be used to evaluate the relative importance of the variables to the design process. Figure 7 briefly summarizes the relative energy requirements estimated in this way.

A 20 percent error in selection of the design temperature is not nearly as serious as a similar magnitude error in estimating organic loading or the hydraulic loading to the aeration tank. A 20 percent error in estimating the SRT also has considerably less impact.

The most significant impact on the estimated energy requirement was caused by the $\beta$ value, assuming a 20 percent lower value. The importance of $\alpha$ was similar to that of the organic and hydraulic loadings.

The serious question which arises from this type of analysis is, what is the potential for error in predicting the future characteristics of wastewater. The characteristics of domestic wastewater historically have been the easiest to predict. Water consumption and dietary habits have slowly changed in a predictable fashion. However, the increasing cost of energy and the related increased cost of all utilities will slowly bring about significant changes in attitudes about conservation. As this occurs, wastewater characteristics will change. Wastewater strength can be expected to increase. This may carry with it a lowering of $\alpha$ and $\beta$ values. Such an occurrence may mean a number of closely designed plants will find aeration system deficiencies as they approach design life and/or design hydraulic loading.

Industrial waste characteristics are predictable as long as process operations remain relatively static. Industry is much more sensitive to economic benefits of process adjustment through conservation if process performance can be assured. The trend toward increased conservation in industry is well documented. This trend will require careful evaluation on a site by site basis for selection of proper design parameters.

## CONCLUSIONS

The design of aeration systems in North America can follow a variety of approaches. System sizing relationships based on experience and on the rationally developed system models are used. The trend is toward improved

Table 4 . Condition and Relationships used to Evaluate Design Sensitivity

a. Reference Values Used

| Parameter | Value | Units |
|---|---|---|
| Influent $BOD_5$, $S_0$ | 250 | mg/L |
| Temperature | 20 | °C |
| Temperature corrections | | |
| Biological rates | 1.02 | - |
| Aeration rates | 1.024 | - |
| Metabolism rate constant, $k_m$* | 15.0 | 1/h |
| Synthesis rate constant, $k_s$* | 10.4 | 1/h |
| Endogenous rate constant, $k_e$* | 0.01 | 1/h |
| Aeration time, t | 5 | h |
| Solid retention time, $\Theta_c$ | 120 | h |
| Saturation $O_2$ concentration, $C_S$ | 9.1 | mg/L |
| Mixed liquor $O_2$ concentration, $C_L$ | 2.0 | mg/L |

b. Adjusted Values Used (20% error)

| Parameter | Values | | Units |
|---|---|---|---|
| | Lower | Upper | |
| $S_0$ | 200 | 300 | mg/L |
| Temperature | 16 | 24 | °C |
| $k_m$ | 13.85 | 16.24 | 1/h |
| $k_s$ | 9.61 | 11.26 | 1/h |
| $k_e$ | 0.0092 | 0.0108 | 1/h |
| t | 4 | 6 | h |
| $\Theta_c$ | 9.6 | 144 | h |

c. Equations and Relationships Used

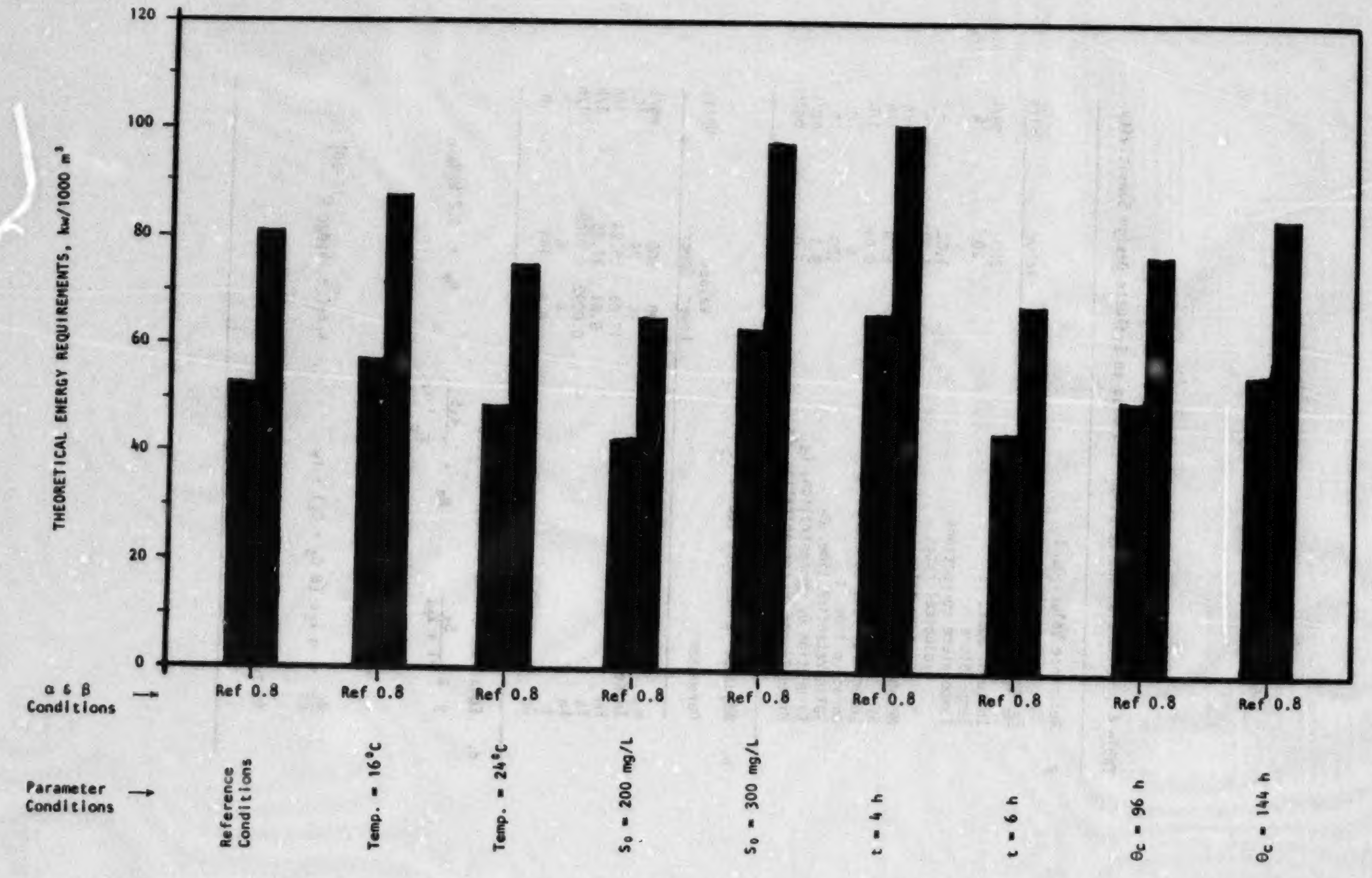
$$S = \frac{S_0}{1 + k_m t} \qquad M_a = \frac{k_s S}{\frac{1}{\Theta_c} + k_e} \qquad M_e = 0.2\, k_e M_a \Theta_c$$

$$r_r = \frac{1.5\,(S_0 - S)}{t} - \frac{1.42\,(M_a + M_e)}{\Theta_c}$$

$$\frac{dc}{dt} = \alpha\, k_L a\,(\beta\, C_s - C_L) - r_r \qquad k_L a_T = k_L a_{20}\, \Theta^{(T-20)}$$

$$O_2 \text{ Transfer Efficiency} = 1.5\, \frac{\text{kg } O_2}{k_w\text{-h}}$$

Figure 7: Results of parameter analysis for rational design of aeration systems.



understanding of the requirements of the biological systems and the aeration system. There is still a major problem with translating the results of testing an aeration system in one tank configuration and size to another.

Review of design parameters and data requirements for sizing aeration systems shows that organic and hydraulic loading parameters are of major importance. However, the β factor and the $C_L$ value to be maintained in the aeration tank are the most significant parameters to the design.

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## DISCLAIMER

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# DESIGN PROTOCOL FOR AERATION SYSTEMS - UK PERSPECTIVE

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## ABSTRACT

The successful design of an activated-sludge aeration system depends upon calculating the mass of oxygen that must be supplied to the process in order to produce an effluent of specified quality. Maximum aeration efficiency is achieved by taking into account the variations in oxygen demand that will occur and designing the aeration system to meet these requirements.

Factors that must be considered include the effects of aeration tank geometry and aerator configuration on the performance of the overall aeration system. In order to achieve optimum aeration efficiency it is also necessary to match the supply of oxygen with demand such that excessive dissolved oxygen concentrations do not occur within the aeration tank.

Recent surveys of aeration systems have shown that potentially the fine-bubble diffused-air system is capable of achieving high aeration efficiencies. Details of process modifications carried out at a large sewage treatment works in the UK to enhance the aeration efficiency of a fine-bubble system are given in an appendix.

## 1. INTRODUCTION

The rate of treatment of waste water by the activated-sludge process is limited by the rate at which micro-organisms in the sludge can biochemically oxidize the impurities, it cannot be increased simply because dissolved oxygen (DO) is readily available. The maximum rate will depend on a number of factors including treatability of the waste water, viability of micro-organisms in the sludge, temperature of aerated sludge, and concentration of dissolved oxygen which should be maintained above critical values. For carbonaceous oxidation the rate-limiting concentration is normally about 0.5 mg/l, and for nitrification the optimal concentration is about 2 mg/l.

The rate of treatment will also be affected by the need to produce consistently an effluent of required quality.

Having established the BOD loading rate of the sludge to produce an effluent of required BOD, with or without nitrification, it is then possible to calculate the concentration of suspended solids that should be maintained in the aeration tank and the period of aeration. From these values the wastage rate of sludge can be calculated and from assumptions about the settleability of the sludge the size of final settlement tank may be estimated.

The total oxygen requirement can be calculated from the BOD to be satisfied and the ammoniacal nitrogen to be oxidised. The aeration intensity (g $O_2/m^3$ of aeration tank per hour) can then be estimated and the aeration system designed to provide the required intensity at expected loading rates which will vary diurnally and maybe seasonally.

In this paper the method adopted in the UK for design of an aeration system is described, taking into account the factors outlined above in order to achieve minimum use of energy for aeration and hence maximum aeration efficiency.

## 2. OXYGEN REQUIREMENTS

### 2.1. Total Demand

The total rate of oxygen consumption by the micro-organisms in an aeration tank is equal to the rate oxygen is required for carbonaceous oxidation by heterotrophic bacteria, plus the rate of oxidation of ammoniacal nitrogen by autotrophic bacteria, less the rate at which nitrate is denitrified for carbonaceous oxidation.

To satisfy BOD, the heterotrophic bacteria consume oxygen for their growth and for respiring endogeneously. To achieve full nitrification, the additional amount of dissolved oxygen consumed is 4.3 times the concentration of ammoniacal nitrogen oxidised to nitrate (assuming 6% of the nitrogen is assimilated by the bacteria). Where anoxic conditions exist in an aeration tank, by design or because of inadequate aeration, the amount of oxygen available to satisfy BOD is equal to 2.83 times the concentration of nitrate-nitrogen denitrified(1).

Experience in the UK and the USA(2,3) has indicated that the total amount of oxygen required for treatment of settled sewage can be calculated from the following empirical equation.

$$R = 0.75\ F\ (C_{BS}-C_{BE}) + 2 \times 10^{-3}\ C_M\ V + 4.3\ (C_{ANS}-C_{ANE})F \qquad (1)$$
$$- 2.83 \left[(C_{ANS}-C_{ANE}) - C_{ONE}\ F\right]$$

where R is the mass of oxygen required (g/h)

F is the flow rate of settled sewage ($m^3/h$)

$C_{BS}$ is the BOD of settled sewage (mg/l)

$C_{BE}$ is the BOD of effluent (mg/l)

$C_M$ is the concentration of suspended solids in the aeration tank (mg/l)

$C_M$ is the concentration of suspended solids in the aeration tank (mg/l)

$V$ is the volume of the aeration tank ($m^3$)

$C_{ANS}$ is the ammoniacal-N content of settled sewage (mg/l)

$C_{ANE}$ is the ammoniacal-N content of the effluent (mg/l)

and $C_{ONE}$ is the nitrate-N content of the effluent (mg/l)

If nitrite were formed as a significant proportion of the oxidised nitrogen in the effluent, the amount of oxygen required for nitrification and available from denitrification would be reduced accordingly. However, it is unlikely that nitrite will constitute a high proportion of the oxidised nitrogen in an effluent unless the waste water contains an inhibitor to the autotrophic bacteria which oxidise nitrite to nitrate. Any nitrate or nitrite present in the settled sewage which is reduced to nitrogen by anoxic conditions in the aeration tank will also reduce the amount of oxygen required for treatment.

The above equation can be simplified, so that

$$R = 0.75\ F\ (C_{BS}-C_{BE}) + 2 \times 10^{-3}\ C_M\ V + 1.47\ F\ (C_{ANS}-C_{ANE}) + 2.83\ F\ C_{ONE} \qquad (2)$$

When treating screened and degritted raw sewage, it has been found(4) that the amount of oxygen required is greater than that calculated from the above equation. However, by changing the first term from 0.75F to 1.0F the oxygen used for such treatment can be calculated with reasonable accuracy using the modified equation.

Data have been obtained which indicate that the above equations can be used to predict the oxygen demand of plants operated within the range of sludge loadings from about 0.05 to 1 g BOD/g sludge suspended solids day.

## 2.2. Variations in Demand

The total quantity of oxygen will vary depending on the mass of sludge in the aeration tank ($C_M V$), which will vary with the growth rate of the micro-organisms (and hence the wastage rate of the sludge), which will depend on the sludge loading rate (g BOD/g sludge day), and this will vary with the effluent quality required.

A relation between the BOD of effluent and the loading rate of sludge has been obtained (Fig. 1) with pilot plant treating settled sewage of average BOD 250 mg/l at temperatures of about 15°C(5). This relation has been substantiated by data obtained from full-scale plants treating settled sewage with BOD in the range from 150 to 250 mg/l.

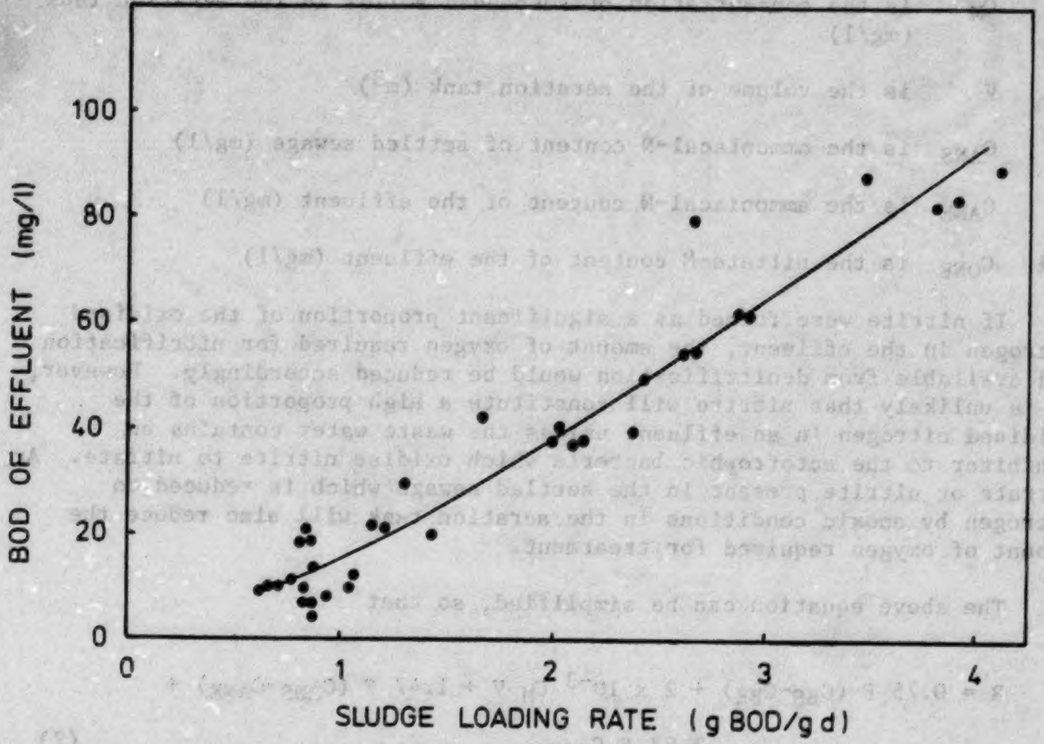

Figure 1. Relation between loading rate of settled domestic sewage (average BOD 250 mg/l and BOD of final effluent

From the curve (Fig. 1) it can be seen that to produce an effluent of BOD of 20 mg/l, the sludge loading should not exceed about 1 g BOD/g sludge. Assuming the settleability of the sludge and the horizontal area of the final settlement tank were sufficient to operate the plant with a mixed liquor suspended solids (MLSS) of 3000 mg/l, the rate of oxygen demand to satisfy the BOD of 230 mg/l (excluding nitrification) would be about 92 $g/m^3$ h. This calculation is based on substitution of the appropriate data into the first two terms in Equation 1 which relate to carbonaceous oxidation. Thus the rate of oxygen demand ($g/m^3$ h) is equal to

$$\frac{0.75(250-20) + 2 \times 10^{-3} \times 3000 \times 2}{2} \text{ or } 92, \ g/m^3 \ h$$

where V/F has the value of 2 h.

If nitrification were required and the maximum loading rate to ensure the growth of nitrifying bacteria was assumed to be 0.2 g/g d, the rate of oxygen demand to oxidise ammoniacal nitrogen and satisfy BOD could be calculated from the first three terms of Equation 1. Assuming the ammoniacal N to be oxidised to nitrate was 30 mg/l and the MLSS were 3000 mg/1, the rate of demand would be equal to

$$\frac{0.75(250-10) + 2 \times 10^{-3} \times 3000 \times 10 + 4.3 \times 30}{10} \ g/m^3 \ h,$$

or 37 $g/m^3$ h, where V/F has the value of 10 h.

If an anoxic zone were provided to denitrify the nitrate in the recycled sludge and the rate of sludge recycle was equal to the flow-rate of sewage so that 50% of the nitrate was reduced, the total rate of demand for oxygen calculated from Equation 2 (with the same values for the other conditions assumed for the nitrifying plant) would be equal to

$$\frac{0.75 \ (250-10) + 2 \times 10^{-3} \times 3000 \times 10 + 1.47 \times 30 + 2.83 \times 15}{10} \ g/m^3 \ h$$

or 32.6 $g/m^3$ h, where V/F has the value of 10 h.

Diurnal variations in rate of flow of waste water and in concentrations of BOD to be satisfied and ammoniacal nitrogen to be oxidised will result in significant variations in oxygen demand. These variations can be calculated using Equation 2 provided the relevant data were available. In practice, it has been found that, to avoid lack of dissolved oxygen limiting the rate of treatment, the period for which the total demand should be calculated is equal to the retention time of mixed liquor (recycled sludge plus sewage) in the aeration tank. The demand thus calculated could be used to determine the rate of oxygen transfer required from the aeration system in a 'uniformly-mixed' aeration tank. However, most activated-sludge plants have aeration tanks which have 'plug-flow' mixing characteristics in order to produce sludges of good settleability.

In 'plug-flow' aeration tanks the demand for oxygen will be greater at the inlet end of the aeration tank where the availability of substrate does not limit the rate of oxidation compared with that at the outlet end where the rate of oxidation is limited because most of the BOD will have been satisfied. In order to estimate the variation in demand along the length of a 'plug-flow' aeration tank it is necessary to have some information about the treatability or biodegradability of the sewage as the presence of substances which are potentially inhibitory to biochemical oxidation will have an effect. In the absence of such information it would have to be assumed that data obtained from the treatment of sewage which contain low proportions of industrial effluent could be applied.

In the UK, data have been obtained from three large treatment works(6) and from pilot-scale studies(7) which show that the rate of oxygen demand for nitrification is constant from the inlet to the outlet of a 'plug-flow' aeration tank provided the concentration of dissolved oxygen is always above about 2 mg/1 and the concentration of ammoniacal nitrogen has not reached the rate-limiting value of about 0.5 mg/l.

To satisfy BOD, the demand at the inlet will depend on dilution rate with recycled sludge and with the contents of the aeration unit which will depend on the dispersion rate (and hence the degree of 'plug-flow') of the mixed liquor. Little information has been published to enable accurate calculation to be made of the effect of such variables on the demand for oxygen. Observations at a number of treatment works in the UK, having aeration tanks in which there exists 'plug-flow' because there is a high length to width ratio (>20) and the dispersion number(8) is very low (about 0.01 to 0.1), indicate that on average about 60% of the BOD was satisfied in the first 20% of the aeration tank volume. In the remaining volume of the aeration tank about 15% of the BOD was satisfied in the next 20% of the aeration tank volume followed by 10, 10, 5% for each following 20% of the volume. It was also found that these values varied diurnally as can be seen from the figures in Table 1 which indicate the range of oxygen demand for each zone.

Table 1. Variation in proportion of oxygen demand along the length of a plug-flow aeration tank

| Proportion of aeration tank volume (%) | Proportion of oxygen demand for carbonaceous oxidation (%) | |
|---|---|---|
| | Average | Range diurnally |
| 20 | 60 | 40-85 |
| 20 | 15 | 5-20 |
| 20 | 10 | 5-15 |
| 20 | 10 | 5-15 |
| 20 | 5 | <1-10 |

If the results shown in Table 1 were related to the examples calculated previously for a non-nitrified effluent and a fully nitrified effluent, without and with denitrification, the following distribution of oxygen demands within the aeration tank can be calculated.

1. Non-nitrified effluent (BOD <20 mg/l)

| | |
|---|---|
| BOD of settled sewage | 250 mg/l |
| MLSS | 3000 mg/l |
| Sludge loading | 1 g/g d |
| Period of aeration of sewage | 2 h |
| Total oxygen demand | 184 g/m$^3$ |

| Proportion of tank volume (%) | | Rate of oxygen demand in each portion of tank (g/m$^3$ h) | |
|---|---|---|---|
| | | Average | Range |
| Inlet | 20 | 276 | 184-391 |
| | 20 | 69 | 23-92 |
| | 20 | 46 | 23-69 |
| | 20 | 46 | 23-69 |
| Outlet | 20 | 23 | <5-46 |

## 2. Nitrified effluent (BOD <20 mg/1 Ammoniacal N <1 mg/1)

| | |
|---|---|
| BOD of settled sewage | 250 mg/1 |
| Ammoniacal nitrogen of settled sewage | 30 mg/1 |
| MLSS | 3000 mg/1 |
| Sludge loading | 0.2 g/g d |
| Period of aeration of sewage | 10 h |
| Total oxygen demand | 370 g/m$^3$ |
| Oxygen demand for nitrification | 129 g/m$^3$ |
| Oxygen demand for carbonaceous oxidation | 241 g/m$^3$ |

If conditions in the aeration tank (DO and concentration of ammoniacal N) were such that the rate of oxygen demand for nitrification were constant throughout the period of aeration, the rate of demand in each portion of the tank would be 12.9 g/m$^3$ h. It can be assumed that the rates of oxygen demand for carbonaceous oxidation would again be divided in the proportions shown in Table 1 so that the rates can then be calculated and they are given below.

| Proportion of tank volume (%) | | Rate of oxygen demand in each portion of tank ($g/m^3$ h) | |
|---|---|---|---|
| | | Average | Range |
| Inlet | 20 | 72 | 48-102 |
| | 20 | 18 | 6-24 |
| | 20 | 12 | 6-18 |
| | 20 | 12 | 6-18 |
| Outlet | 20 | 6 | <1-12 |

The total rates of demand for oxygen (to satisfy BOD and oxidise ammonia) in the aeration tank would be as follows.

| Proportion of tank volume (%) | | Rate of oxygen demand in each portion of tank ($g/m^3$ h) | |
|---|---|---|---|
| | | Average | Range |
| Inlet | 20 | 85 | 61-115 |
| | 20 | 31 | 19-37 |
| | 20 | 25 | 19-31 |
| | 20 | 25 | 19-31 |
| Outlet | 20 | 19 | 13-25 |

### 2.3. Nitrified effluent with denitrification (BOD <20 mg/l Ammoniacal N <1 mg/l Nitrate N <15 mg/l)

In this case, the 15 mg/l of nitrate N that would be used to satisfy BOD would reduce the oxygen demand for carbonaceous oxidation from 241 $g/m^3$ to 197 $g/m^3$. This would reduce the rates of demand in each portion of the tank although, like the example above, the rate of oxygen demand for nitrification would remain constant at 12.9 $g/m^3$ h.

## 3. SUPPLY OF OXYGEN

In most treatment plants in the UK, dissolved oxygen is supplied to the activated-sludge process by either fine-bubble diffused-air or mechanical surface aeration. There are 6 oxygenated plants using the Unox process, about 10 with oxygen injection using the Vitox process to uprate existing aeration equipment and 2 Deep-Shaft plants.

Air is normally diffused through ceramic diffusers into the mixed liquor near to the bottom of the tank to produce fine-bubbles (2 to 5 mm diameter) so as to achieve adequate mixing and maximum contact period.

There are two basic types of mechanical surface device, one which has a vertically mounted (cone-type) aerator and the other a horizontally mounted (brush-type) aerator. Tests in the UK of these aeration devices have shown(9) that the fine-bubble diffused-air system can have the greatest aeration intensity (g $O_2$ supplied/$m^3$ aeration tank capacity per hour) and efficiency (g $O_2$ supplied/Wh), provided the system is correctly installed, operated and maintained (Table 2).

The aeration intensity and aeration efficiency are directly affected by changes in the deficit of dissolved oxygen. Thus an aeration device operating in mixed liquor to maintain the concentration of dissolved oxygen at 20% of the air-saturation concentration will have only 80% of the intensity and efficiency of oxygen transfer shown in Table 2.

Table 2. Aeration intensity and efficiency of conventional systems of aeration as measured in tap water (2 to 8 m deep) containing surfactant (about 5 mg/l). Similar results have been obtained in aerated mixed liquor

| Aeration system | Maximum aeration intensity ($g/m^3$ h) | Aeration efficiency (kg/kWh) |
|---|---|---|
| Diffused-air | | |
| Fine-bubble (<5mm diameter) | 200 | 1.5-3.6 |
| Coarse-bubble (>5 mm diameter) | 100 | 0.9-1.2 |
| Mechanical surface aerators | | |
| Rotating vertically | 125 | 1.5-2.2 |
| Rotating horizontally | 100 | 1.2-2.4 |

All values obtained at maximum deficit of dissolved oxygen
Aeration efficiency calculated from total energy supplied.

Comparison of the aeration intensities given in Table 2 with the examples calculated in Section 2.2 shows that the rates of oxygen transfer by the aerators could limit the rate of treatment at the inlet end of 'plug-flow' aeration tanks particularly with mechanical surface aerators.

The methods used in the UK of testing an aeration system to determine its ability to transfer oxygen into solution have been described(9). The results obtained have shown(9) that conducting tests in 'clean water' using the unsteady-state method is not entirely satisfactory because 'clean water' can easily be contaminated by traces of surface-active substances (including some that might be added with chemicals used to deoxygenate the water). It is now widely accepted in the UK that one or two tests should be duplicated with the second test conducted in water to which anionic surface-active agent has been added to achieve an average concentration during the test of about 5 mg/l. The surface-active agent is added to the water used for testing aerator for three important reasons.

1. It minimises the effects of traces of other surface active contaminants (such as oil and greases).

2. It affects the rate of oxygen transfer so that the rate measured is similar to that in mixed liquor at the inlet end of the aeration tank. At the outlet end the rate is usually similar to that in 'clean tap water'.

3. It enables reproducible results to be obtained irrespective of the purity of the water - thus tap water, river water, or final effluent may be used.

To ensure that the rate of supply of oxygen by an aerator is equal to lished over a wide range of conditions. This is because the rate of oxygen transfer is affected by changes in the conditions such as degree of purification of sewage, configuration of aerators in the aeration tank and geometry (volume, depth and shape) of the aeration tank, and variation in concentration of dissolved oxygen. Considerable data are now available in the literature which describe the effects of these changes in conditions on the performance of many different types of aeration device(10-13). In this paper are summarised the effects such changes in conditions have on the performance of an aerator with particular reference to the fine-bubble diffused-air system because most recent research in the UK has concentrated on this system.

### 3.1. Effect of Changes in Degree of Purification

The performance of an aerator can be measured in aerated activated-sludge during treatment of sewage using the steady-state method(9). Such measurements have been used to establish the effects of a surface-active agent as outlined above. As most tests of aerators are carried out in 'clean water' a factor of proportionality, termed 'alpha' is used to relate the performance of an aerator in 'clean water' to that in aerated activated-sludge.

The value of 'alpha' for a given aeration system is equal to the performance of the system (aeration efficiency or intensity) in aerated liquor under specified conditions divided by the corresponding performance under similar conditions in clean water. Published results show that the

**performance of an aeration system varies significantly according to the nature of the liquor, the contaminants present, and the intensity of aeration(11).**

**The value of 'alpha' for a given aerator does not remain constant. The efficiency of transfer of oxygen into aerated liquor varies with the degree of purification of the waste water. Experiments with a fine-bubble aeration system, operating in an essentially 'plug-flow' unit, have shown (Fig. 2)**

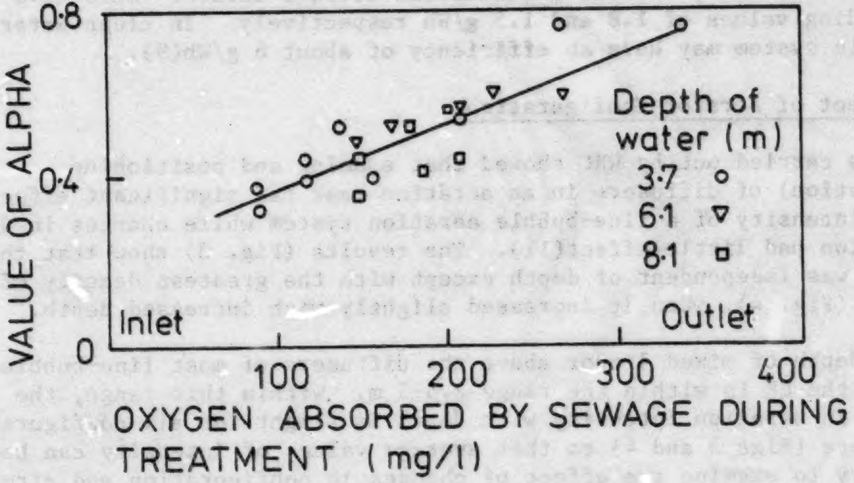

**Figure 2. Variation of alpha factor with degree of purification of sewage**

**variations in 'alpha' from 0.3 at the start of the treatment (when the waste water was first brought into contact with recycled sludge) to 0.8 at completion of treatment (when a high-quality, fully nitrified effluent had been produced). The results showed that the performance of the aerator at the inlet end of the plug-flow unit was similar to that measured in tap water with 5 mg of added anionic surfactant per litre, under similar conditions of air-flow rate, temperature, tank geometry and diffuser configuration. The performance at the outlet of a plug-flow unit corresponds approximately to that measured in clean water under similar conditions.**

**The performance of mechanical surface aerators installed in a plug-flow aeration unit also changes as the sewage becomes purified. With such aerators the value of 'alpha' may be about 1.2 initially and may decrease to about 1, corresponding to a decrease in the concentration of contaminants. However, in some circumstances, particularly with hori-zonatally rotating aerators, the initial value of 'alpha' may be about 0.8 and it may increase to 1 as the concentration of contaminants decreases.**

'Alpha' values do not indicate the relative efficiency of aeration systems. A fine-bubble diffused-air system may have an 'alpha' value as low as 0.4 at the inlet end of a plug-flow aeration unit, and in a mechanically operated system it may be as high as 1.2. The relative efficiencies of different aerators can be judged only by comparison of their mass-transfer coefficients, aeration intensities, or aeration efficiencies when tested under similar conditions in aerated liquor (or in tap water with or without added surfactant). For example, a fine-bubble aeration system may have an aeration efficiency of 2 g/Wh at the inlet of a plug-flow aeration tank and 4 g/Wh at the outlet, whereas a mechanical surface aerator could have corresponding values of 1.8 and 1.5 g/Wh respectively. In clean water the fine-bubble system may have an efficiency of about 6 g/Wh(9).

### 3.2. Effect of Aerator Configuration

Tests carried out by WRC showed that spacing and positioning (configuration) of diffusers in an aeration tank had significant effect on aeration intensity of a fine-bubble aeration system while changes in depth of immersion had little effect(11). The results (Fig. 3) show that the intensity was independent of depth except with the greatest density of diffusers (Fig. 4), when it increased slightly with increased depth.

The depth of mixed liquor above the diffusers of most fine-bubble plants in the UK is within the range 2.5-5 m. Within this range, the variation in aeration intensity with depth is slight for all configurations of diffusers (Figs 3 and 4) so that average values of intensity can be used confidently to examine the effect of changes in configuration and air-flow rates (Table 3).

From these results (Table 3) the following assumptions can be made within the limits of the data presented.

1. For a given flow rate of air per diffuser, the change in aeration intensity is proportional to the change in number of diffusers/$m^2$ surface area.
2. For a given number of diffusers/$m^2$ surface area, the aeration intensity changes in proportion to changes in flow-rate of air/dome.

The aeration intensity of surface aeration systems has been shown to vary with rotation rate and depth of immersion of the aerator and with geometry (depth and surface area) of the aeration tank(10,13).

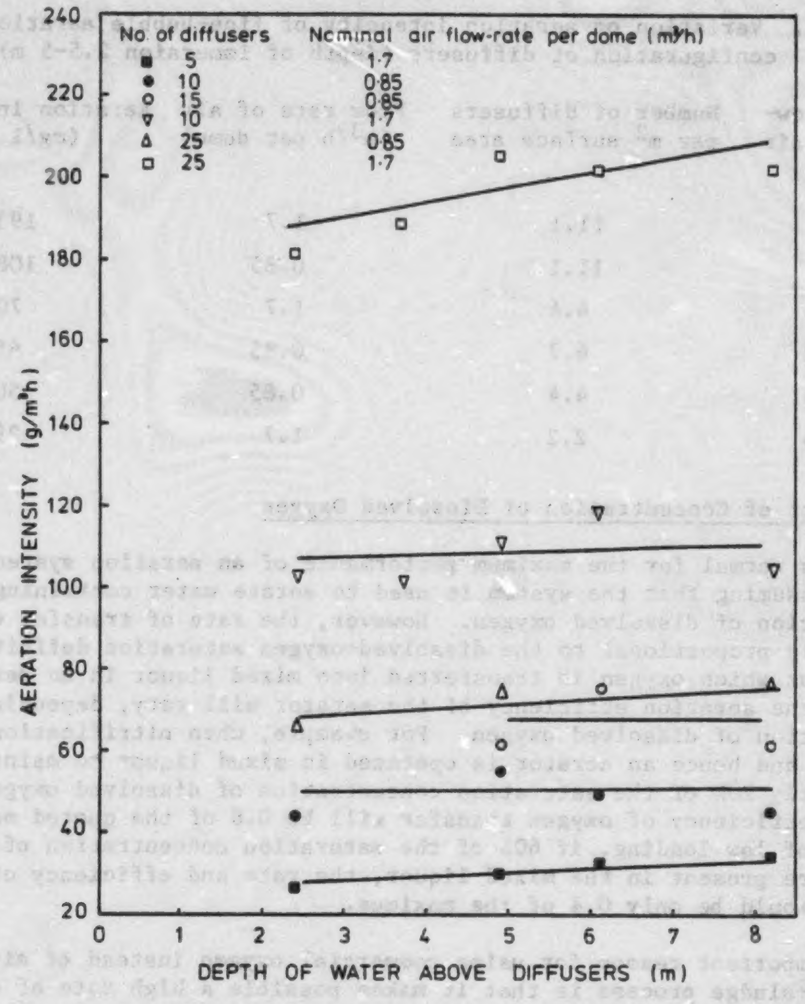

**Figure 3.** Variation of aeration intensity with depth of water for different configurations of diffusers (as shown in Fig.4) and air flows when aerating tap water containing surfactant (about 5 mg/l) with standard dome diffusers (178 mm diameter) sited on the bottom of the tank.

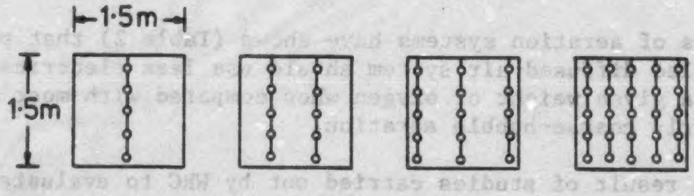

**Figure 4.** Configuration of diffusers in aeration tank (plan not to scale)

Table 3. Variation of aeration intensity of fine-bubble aeration with configuration of diffusers (depth of immersion 2.5-5 m)

| Total flow-rate of air ($m^3$/h) | Number of diffusers per $m^2$ surface area | Flow rate of air ($m^3$/h per dome) | Aeration intensity (mg/l h) |
|---|---|---|---|
| 42 | 11.1 | 1.7 | 193 |
| 21 | 11.1 | 0.85 | 108 |
| 8.4 | 4.4 | 1.7 | 70 |
| 25.2 | 6.7 | 0.85 | 65 |
| 8.4 | 4.4 | 0.85 | 50 |
| 8.4 | 2.2 | 1.7 | 29 |

### 3.3. Effect of Concentration of Dissolved Oxygen

It is normal for the maximum performance of an aeration system to be quoted assuming that the system is used to aerate water containing zero concentration of dissolved oxygen. However, the rate of transfer of oxygen is directly proportional to the dissolved-oxygen saturation deficit. Hence the rate at which oxygen is transferred into mixed liquor in an aeration tank and the aeration efficiency of the aerator will vary, depending on the concentration of dissolved oxygen. For example, when nitrification is required, and hence an aerator is operated in mixed liquor to maintain consistently 20% of the saturation concentration of dissolved oxygen, the rate and efficiency of oxygen transfer will be 0.8 of the quoted maximum. At times of low loading, if 60% of the saturation concentration of dissolved oxygen were present in the mixed liquor, the rate and efficiency of oxygen transfer would be only 0.4 of the maximum.

An important reason for using commercial oxygen instead of air in the activated-sludge process is that it makes possible a high rate of oxygen transfer when operating with high concentrations of dissolved oxygen (40-60% of the air saturation value) in the mixed liquor. However, operation of an activated-sludge plant at such high concentrations of dissolved oxygen has little effect on the rate of treatment, and hence the size of the plant, or the quality of effluent obtained(15).

## 4. OPTIMISATION OF AERATION EFFICIENCY

Tests of aeration systems have shown (Table 2) that potentially the fine-bubbled diffused-air system should use less electrical energy to dissolve a given weight of oxygen when compared with most other systems, particularly coarse-bubble aeration.

As a result of studies carried out by WRC to evaluate a fine-bubble aeration system(9,11), and further work undertaken by the US EPA in co-operation with the WRC(3), Thames Water Authority, Severn-Trent WaterAuthority and others, the following factors have been determined as important in order to obtain optimum performance (lowest energy input).

(i) The geometry of the aeration basin and configuration of the diffusers are probably the most significant factors to affect aeration efficiency. It is recognised that there are certain advantages to plug-flow of liquor along the aeration tank, particularly in relation to improved settleability of activated sludge(8), but a length-to-width ratio not exceeding 12:1 could give optimal aeration efficiency. Tapering of aeration intensity by varying the configuration of diffusers along the tank should be provided but not to such an extent that, at times of low BOD and ammoniacal nitrogen load, the required flow-rate of air is below the minimum specified by the manufacturers or the intensity of mixing is inadequate to keep the activated sludge in suspension. The optimum depth of liquor in the aeration tank is probably between 4.5 and 6 m(3).

(ii) Excessively high concentrations of dissolved oxygen (DO) have little benefit to the treatment process but significantly reduce the aeration efficiency. Monitoring and control of DO by varying the air-flow rate should result in an increase in aeration efficiency and a saving of energy. To achieve effective control of DO, a minimum of 3 independently controlled air-grids, each with air-metering and linked to a DO probe via a programmable controller, would be required. The control system could be provided by a mini-computer capable of varying the output of air supplied by a variable number of blowers one of which could be driven by a speed-controlled DC electric motor.

(iii) The intensity of aeration should correspond to the requirement for oxygen (Para. 2.2). Changes in temperature and treatability of waste water will affect the rate of biochemical oxidation so that the maximum intensity of aeration should be based on the maximum rate of carbonaceous and nitrogenous oxidation anticipated, preferably as determined from experience rather than based on rigid application of such data as are presented in this paper.

## 5. DISCUSSION

It is probable that the saving of energy in the activated-sludge process requires additional capital expenditure. The benefits can only then be judged by economic evaluations, taking into account any differences in capital and operating costs. Such costs are beyond the scope of this paper, and data, such as capital costs, are best obtained from quotations of manufacturers in specific circumstances.

Experience in the UK, over many years, has shown that most aeration systems operated reliably, have few significant problems, and are not expensive to maintain. Contrary to popular belief regarding clogging of fine-bubble diffusers, the data obtained from a survey of 19 full-scale sewage-treatment plants, the smallest treating 0.15 $m^3/s$ and the largest treating 9 $m^3/s$, showed that the frequency of cleaning diffusers ranged from once every 4 years to once in 15 years; many of the plants had been in operation for 6 to 12 years without the need to clean the diffusers(3). Facilities were provided at all the works to clean the air supplied to the diffusers, and their capital and operating costs were relatively small compared with total costs of installation and operation of the aeration system.

It was unfortunate that many of the plants surveyed(3) were being operated at aeration efficiencies significantly lower than the maximum anticipated from pilot and full-scale testing of the fine-bubble aerators(9,11,12). In most cases, lack of facilities to control air-flow rates resulted in inefficient operation of the aerators, but at a few works operation at sludge-loading lower than design, and/or the adverse effects of industrial waste waters, were thought to influence aerator performance. All the plants produced effluents of a high quality, most with oxidation of ammonia and some with controlled denitrification in anoxic zones. The average oxygen-transfer efficiency (calculated to correspond to the rate of oxygen transfer at a DO of 20% of the air-saturation concentration) ranged from 2.9 to 1.0 g/Wh, with a median value of 2.2 g/Wh. At maximum deficit of dissolved oxygen (corresponding to operation at zero DO) the median value of efficiency would be 2.75 g/Wh, which is in good agreement with the results obtained in water plus 5 mg anionic surfactant/l and in aerated liquor(9). At the works surveyed, the observed aeration efficiencies could be increased by changes in the operation and control of the aeration equipment which could save about a third of the energy previously used.

The energy needed to dissolve oxygen can be minimised by the use of fine-bubble aeration, but there are other factors that should be taken into account in the selection of an aeration system. These include capital cost and size of plant, settleability and rate of production of activated sludge, and also reliability, flexibility, and durability of the aeration equipment. In respect of some of these factors there could be advantages in the use of new processes such as the Deep-Shaft and systems for injection of oxygen which can be used to supplement the aeration intensity provided by existing equipment, particularly during periods of high loading(15). However, to produce an effluent of low BOD, the rate of treatment will be be low and hence there is no need to supply oxygen at a high intensity, except perhaps at the inlet of a 'plug-flow' aeration tank.

Overloading at the inlet of a 'plug-flow' aeration plant could be avoided by making changes to the conditions under which sewage and recycled sludge are brought into contact. For example, such changes could include the provision of a separate tank at the inlet which would be about 15% of the total aeration volume. In the case of a nitrifying plant, this tank should be mixed by sub-surface mechanical stirrers (using energy of about 10 $w/m^3$) to achieve controlled anoxic conditions so that nitrate, available from the recycled sludge fed to the tank; this would reduce the total amount of air required to treat the sewage and may also improve the settleability of the activated sludge(8). In the case of a non-nitrifying plant, the first tank could be used to aerate recycled activated sludge, before the addition of sewage to the second aeration tank. This would minimise the BOD loading at the inlet of the second tank, and reduce the overall sludge loading of the plant for a given concentration of suspended solids in the aerated liquor, so reducing the risk of slime formation on the diffusers and improving the effectiveness of DO control.

To demonstrate the low use of energy, consistent quality of effluent, and reliability, flexibility and cost-effectiveness of using a fine-bubble diffused-air system the Thames Water Authority (TWA) and WRC have set up a demonstration at Rye Meads Sewage Treatment Works. This demonstration is financed by WRC at a total cost of about £500 000 with contributions from the UK Department of Energy (£145 000), United States-EPA ($72 000) and Environment Canada ($60 000).

At Rye Meads two aeration channels have been modified, one to be operated to produce a fully-nitrified effluent (with partial denitrification in an anoxic zone) and the other to produce a non-nitrified effluent of 30:20 (SS:BOD) standard for 95% of the time, and their performance compared (particularly load treated and energy used) with an existing channel which produces a fully-nitrified effluent (also with partial denitrification).

The design protocol described in this paper and the philosophy of design and operation outlined below were used to provide the basis for modifications to the aeration units at Rye Meads. Details of the process modifications carried out at Rye Meads to enhance the aeration efficiency are described in the Appendix.

## 6. SUMMARY

The philosophy of design and operation to achieve maximum aeration efficiency can be summarised as follows.

1. The rate of oxygen supply has to equal the rate of biochemical oxygen demand.

2. To obtain the correct rate of supply requires the correct configuration of diffusers and design of aeration tank geometry.

3. To vary the rate of supply to correspond with changes in oxygen demand requires facilities for monitoring and control of dissolved oxygen by varying the flow-rate of air.

4. To reduce the energy used when the air-flow rate is reduced requires a number of blowers, one or two of which would be of controlled and variable speed.

5. There are other factors which must be considered to avoid loss of efficiency and problems with operation and these are listed below.

(i) Minimum flow-rate of air, at minimum demand rate, should never be lower than the minimum rate specified by the manufacturers if problems of clogging of diffusers, inadequate mixing of sludge and sewage, and low aeration efficiency are to be avoided.

(ii) At all flow-rates of air, the valves used for control of flow rate should be near fully open as possible - one valve should always be almost fully open. Control of flow-rate should be largely achieved by varying the pressure via a variable-speed blower.

(iii) Guidelines specified by the manufacturers for maintenance and cleaning of equipment should be fully observed.

(iv) Growths of slimes on diffusers, resulting in inefficient coarse-bubble aeration can be avoided by improved methods of mixing of sewage and recycled sludge at the inlet of the aeration tank to reduce high loading rate of oxygen demand. The provision of anoxic conditions at the inlet of a nitrifying plant, or aeration of recycled sludge for a non-nitrifying plant, would assist to avoid such problems.

(v) Maximum aeration efficiency would be obtained in 'uniformly-mixed' tanks although aeration tanks with 'plug-flow' mixing characteristics tend to produce activated sludge of good settleability. A length-to-width ratio not exceeding 12:1 could give optimal aeration efficiency without poor settleability. The optimum depth of liquid in the aeration tank is probably between 4.5 and 6 m.

## 7. ACKNOWLEDGEMENTS

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This paper is published by permission of Dr J Cuthbert, Director, Process Engineering, Water Research Centre, UK.

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## DISCLAIMER

The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency, and no official endorsement should be inferred.

# APPENDIX

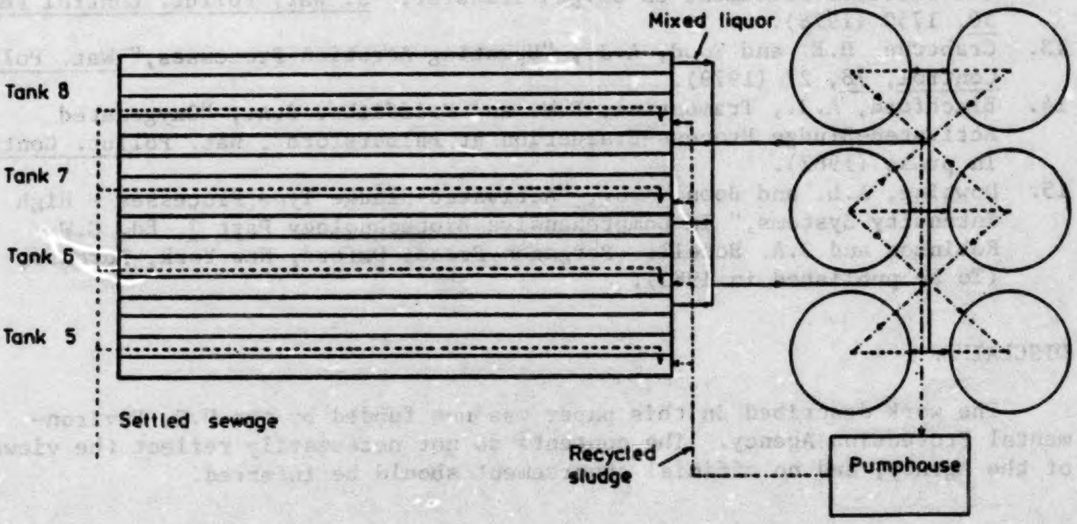
## Details of Process Modifications carried out at Rye Meads STW to enhance aeration efficiency

### A1. INTRODUCTION

There are 12 activated-sludge aeration tanks at Rye Meads arranged as 3 groups, or stages, of 4 tanks. Each stage is associated with a set of six final sedimentation tanks as shown in Figure 1A. Tanks 5 and 8 of Stage 2 were selected for modification for the project. Tank 5 was modified to produce a fully nitrified effluent of 15:10:5 (SS:BOD:Amm N) for 95% of the time and Tank 8 was modified to produce a non-nitrified effluent of 30:20 (SS:BOD) standard for 95% of the time.

Aeration Tank 9 of Stage 3 was selected to be monitored as the control unit.

FIGURE 1A RYEMEADS STAGE 2 LAYOUT OF AERATION AND FINAL TANKS (DIAGRAMMATIC )



### A2. SEPARATION OF PROCESS STREAMS IN TANKS 5 AND 8

In order to make valid comparisons between the two modified aeration tanks it was necessary to isolate them so that each received a separate stream of settled sewage and recycled sludge. Under the previous configuration a single pump was used to return sludge to a common channel at the inlet of Tanks 5-8 inclusive, whence it was distributed as shown in

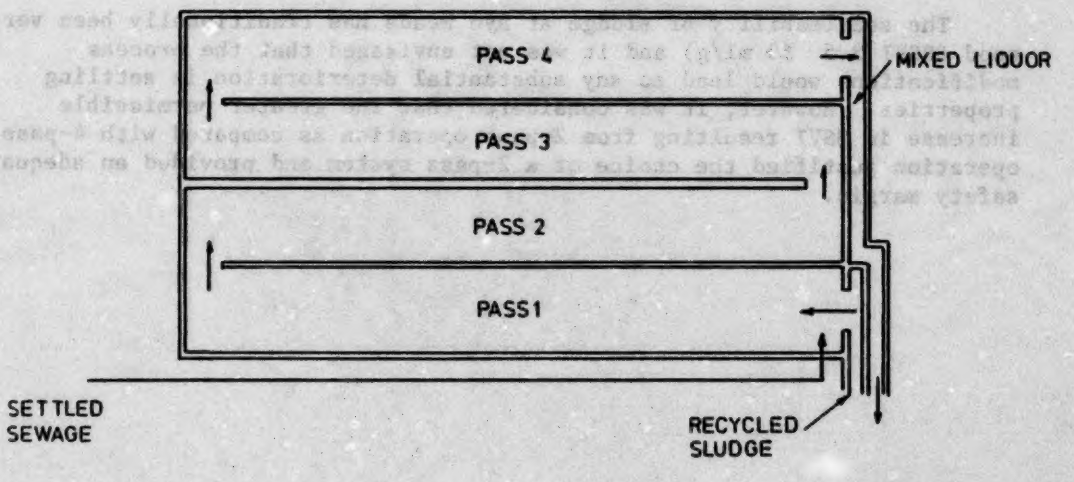
Figure 1A. A second pump was installed to return sludge from Tank 8 to the inlet of that tank only and the existing pump was used to return sludge to Tank 5. Tanks 6 and 7 were drained and sealed for the period of the project and unused mixed liquor and recycled-sludge channels were blanked off. This arrangement allowed one, two or three final sedimentation tanks to be used for the settlement of sludge from each of Tanks 5 and 8. It was also necessary to install an additional recycled-sludge flowmeter, a new recycled-sludge channel and a surplus-sludge metering pump for Tank 8 and a surplus-sludge metering pump for Tank 5.

## A3. REDUCTION OF AERATION TANK LENGTH TO WIDTH RATIO AND PROVISION OF INCREASED SEWAGE FLOW TO AERATION TANKS

Each aeration tank in Stage 2 consisted initially of 4 passes connected in series as shown in Figure 2A. Each pass is 70 m in length, 4.3 m wide and 3.2 m deep. The total volume of each was therefore about 3800 $m^3$.

The length to width ratio of Tanks 5 and 8 was reduced to half its original value by using only the first two passes of each tank. To achieve this the outlet from Pass 2 to Pass 3 of each tank was sealed and new mixed-liquor outlets were cut in the tank walls to allow flow to the settlement tanks. It was not considered that the reduction in length to width ratio would lead to any problems with sludge settleability.

FIGURE 2A RYE MEADS STW AERATION TANK LAYOUT



Each unmodified aeration tank at Rye Meads receives on average about 100 l/s of settled sewage. Although the modified 2-pass aeration tanks are only half of the original volume it was not proposed to decrease the sewage flow proportionally. Since it is known that increasing the rate of treatment leads to potential increases in aeration efficiency and since the amount of oxygen required per unit BOD satisfied decreases with increasing load, it was decided to design initially for an average sewage flow-rate of 85 l/s in the nitrifying plant and 140 l/s in the non-nitrifying plant. This represents an uprating, in terms of flow per unit tank volume, of about 70% and 180% respectively.

The reduction in length to width ratio and the proposed increase in sewage flow-rates were chosen after consideration of possible effects on sludge settleability and the consequence of such effects on operation of final settlement tank. It was felt that the increase in aeration efficiency resulting from the change in length to width ratio would be small in comparison to that resulting from correct configuration of diffusers, the increased sewage flow-rate, and effective control of aeration. However, to obtain a substantial increase in aeration efficiency in an unmodified 4-pass tank would have required an increase in sewage flow to a rate where overloading of the settlement tanks might have occurred. The conversion to 2-pass operation brought about the desired increase in plant loading by reducing the aeration tank volume. The effect of increasing the sewage flow-rate on the operating characteristics of the final sedimentation tanks is summarised in Table 1A. Table 1A shows in parts (a) and (b) the proposed operating conditions for the tanks modified to produce nitrified and non-nitrified effluent. In part (c) of Table 1A it can be seen that in order to achieve the same sludge loading it would be necessary to increase the sewage flow-rate to the values shown. Under such conditions the sludge settleability would have to remain at values less than those shown (as SSVI at 3.5 g/l) otherwise solids overflow with the effluent would occur.

The settleability of sludge at Rye Meads has traditionally been very good (SSVI 3.5 50 ml/g) and it was not envisaged that the process modifications would lead to any substantial deterioration in settling properties. However, it was considered that the greater permissible increase in SSVI resulting from 2-pass operation as compared with 4-pass operation justified the choice of a 2-pass system and provided an adequate safety margin.

### Table 1A. Effects of increased sewage flowrate on operating characteristics of final sedimentation tanks

(a) Modified Nitrifying Plant

| | Number of Final Settlement Tanks | | |
|---|---|---|---|
| | 1 | 2 | 3 |
| Aeration tank volume ($m^3$) | 1900 | 1900 | 1900 |
| Final tank surface area ($m^2$) | 380 | 760 | 1140 |
| Sewage flowrate (l/s) | 85 | 85 | 85 |
| Recycled sludge flowrate (l/s) | 70 | 70 | 70 |
| MLSS (mg/l) | 5000 | 5000 | 5000 |
| Hydraulic residence time (h) | 3.4 | 3.4 | 3.4 |
| Sludge loading ($d^{-1}$) (for BOD influent of 120 mg/l) | 0.1 | 0.1 | 0.1 |
| Max. permissible $SSVI_{3.5}$ before solids loss (ml/g) | 90 | 110 | 130 |

(b) Modified Non-Nitrifying Plant

| | 1 | 2 | 3 |
|---|---|---|---|
| Aeration tank volume ($m^3$) | 1900 | 1900 | 1900 |
| Final tank surface area ($m^2$) | 380 | 760 | 1140 |
| Sewage flowrate (l/s) | 140 | 140 | 140 |
| Recycled sludge flowrate (l/s) | 105 | 105 | 105 |
| MLSS (mg/l) | 3700 | 3700 | 3700 |
| Hydraulic residence time (h) | 2.2 | 2.2 | 2.2 |
| Sludge loading ($d^{-1}$) (BOD influent of 120 mg/l) | 0.2 | 0.2 | 0.2 |
| Max. permissible $SSVI_{3.5}$ before solids loss (ml/g) | 105 | 145 | 165 |

(c) Unmodified 4-pass Aeration Tank

| | 1 | | 2 | | 3 | |
|---|---|---|---|---|---|---|
| | Nitrifying | Non-nitrifying | Nitrifying | Non-nitrifying | Nitrifying | Non-nitrifying |
| Aeration tank volume ($m^3$) | 3800 | 3800 | 3800 | 3800 | 3800 | 3800 |
| Final tank surface area ($m^2$) | 380 | 380 | 760 | 760 | 1140 | 1140 |
| Sewage flowrate (l/s) | 170 | 280 | 170 | 280 | 170 | 280 |
| Recycled sludge flowrate (l/s) | 140 | 210 | 140 | 210 | 140 | 210 |
| MLSS (mg/l) | 5000 | 3700 | 5000 | 3700 | 5000 | 3700 |
| Hydraulic residence time (h) | 3.4 | 2.2 | 3.4 | 2.2 | 3.4 | 2.2 |
| Sludge loading ($d^{-1}$) | 0.1 | 0.2 | 0.1 | 0.2 | 0.1 | 0.2 |
| Maximum permissible $SSVI_{3.5}$ before solids loss (ml/g) | 70* | 90* | 90 | 105 | 105 | 125 |

* Recycled sludge flowrate per unit tank area approaching critical value

## A4. DESIGN AND INSTALLATION OF MODIFIED DOME-DIFFUSER AERATION SYSTEM

### A.4.1. Estimation of Parameters Required for Calculation of Oxygen Demand

The average flow-rates of 85 l/s and 140 l/s were taken as the basis for the calculation of the oxygen requirements of the nitrifying and non-nitrifying plants respectively. Other design parameters were obtained by examination of several years analytical and operational data provided by TWA staff at Rye Meads.

It was found that the ranges of the parameters required were distributed in such a way as to give approximate straight line relations when plotted on probability graph paper. Figure 3A shows, as an example, the relation found for the ammonia nitrogen concentration in settled sewage. From such relations the values shown in Table 2A were obtained.

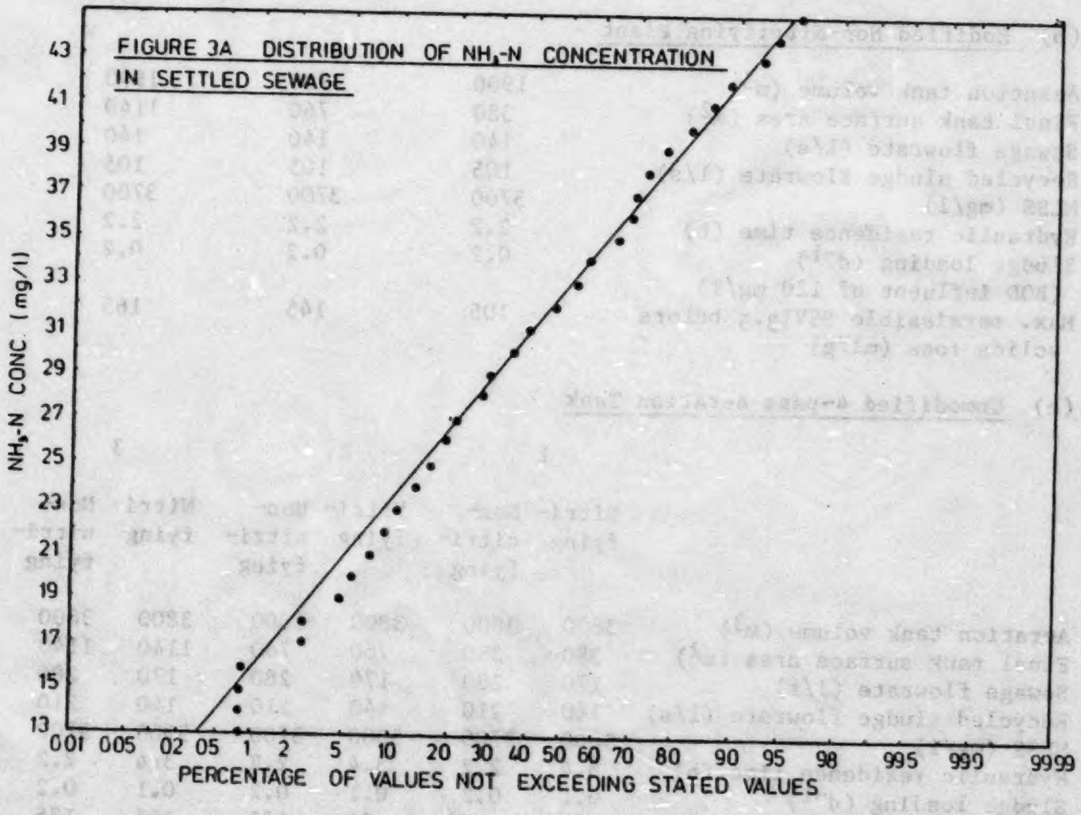


FIGURE 3A DISTRIBUTION OF $NH_4$-N CONCENTRATION IN SETTLED SEWAGE

The range of settled sewage flow-rates shown in Table 2A was calculated by assuming the proposed flow-rates of 85 l/s and 140 l/s were median values. The range of flow-rates expected for these median values was estimated by assuming the flow-rates would vary in the same manner as historical data indicated.

Similarly it was assumed that the modified nitrifying plant would produce a distribution of effluent nitrate nitrogen which was approximately the same as existing aeration tanks.

Table 2A. Parameters used for design of modified aeration systems

| | Nitrifying plant | | | Non-nitrifying plant | | |
|---|---|---|---|---|---|---|
| | 5%< | Median | 95%< | 5%< | Median | 95%< |
| Settled sewage flowrate (l/s) | 64 | 85 | 160 | 100 | 140 | 260 |
| Settled sewage BOD (mg/l) | 75 | 130 | 220 | 75 | 130 | 220 |
| Settled sewage $NH_3$-N (mg/l) | 20 | 33 | 44 | 20 | 33 | 44 |
| Effluent $NO_3$-N (mg/l) | 14 | 22 | 32 | N.A | N.A | N.A |

## A.4.2. Calculation of Oxygen Demand Range in Modified Plants

The range of oxygen demands to be satisfied in the modified aeration tanks was calculated using Equation 2 and the values shown in Table 2A. In addition it was assumed that:-

(i) The MLSS concentrations in the nitrifying plant and the non-nitrifying plant would be maintained, on average, at 5000 mg/l and 3,700 mg/l respectively.

(ii) The BOD in the effluent from the modified plants would be 5 mg/l and 20 mg/l respectively for 95% of the time.

(iii) The effluent ammoniacal N concentration in the nitrifying plant would be less than 1 mg/l or effectively zero.

The calculated range of oxygen demands is shown in Table 3A.

Table 3A. Range of oxygen demands to be satisfied in nitrifying and non-nitrifying aeration tanks

| | Nitrifying | | | Non-nitrifying | | |
|---|---|---|---|---|---|---|
| | 5%< | Median | 95%< | 5%< | Median | 95%< |
| Oxygen demand (kg/d) | 1120 | 1975 | 4910 | 675 | 1320 | 3690 |

### A.4.3. Determination of Number and Arrangement of Dome-Diffusers

The determination of the number of domes required and their arrangement on the floor of the modified aeration tanks was carried out according to the following general procedure, taking into account the criteria mentioned.

(i) The number of domes in each tank should be such that the minimum oxygen demand can be satisfied by a dome operating slightly above the recommended minimum air flow-rate per dome of 0.9 $m^3/h$.

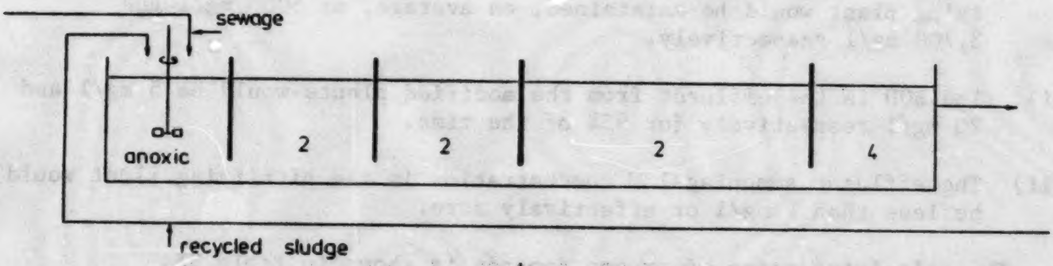
(ii) The maximum oxygen demand should be satisfied by the domes operating with acceptable pressure-drop.

(iii) The arrangements of the domes in the aeration tanks should be tapered from inlet to outlet such that the median oxygen demand is satisfied by the domes operating at average air flow-rate, resulting in the desired uniform dissolved-oxygen concentration throughout the tank.

(iv) The arrangement of domes into separate 'zones' along the length of the aeration tank should facilitate the proposed strategy of dissolved-oxygen control.

Consideration of the above criteria generally results in the calculations required being of an iterative nature and since the criteria are interactive some element of compromise is usually necessary before a reasonable, practical design results.

a, Nitrifying



b, Non-nitrifying

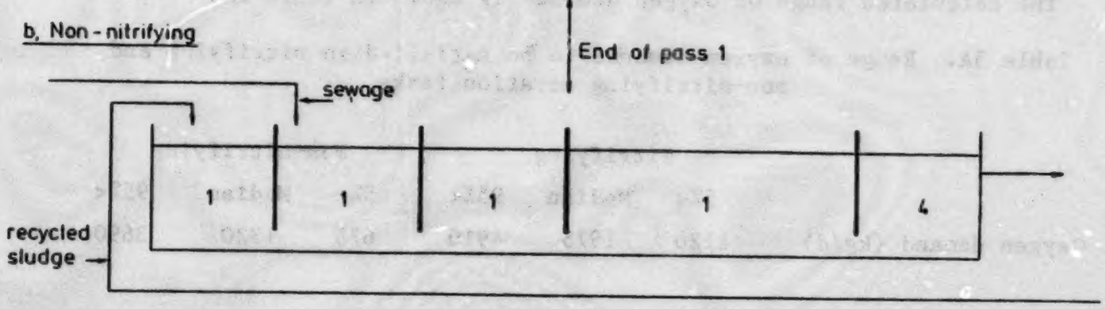


FIGURE 4A PROPOSED D.O. CONCENTRATIONS AND DIVISION OF AERATION TANKS INTO CONTROL ZONES

The basic concept of the layout of the modified tanks at Rye Meads was originally envisaged as shown in Figure 4A. This figure shows the desired dissolved oxygen concentrations considered necessary for efficient operation in the nitrifying and non-nitrifying systems, and the proposed division of the aeration tanks into separate control zones each with its own fully modulating air supply.

In the nitrifying plant an anoxic zone was incorporated as shown at the inlet to the aeration tank. Anoxic zones result in the following process advantages:-

(i) The settleability of sludge in plants incorporating anoxic zones has been shown to be superior to that from similar plants without anoxic zones.

(ii) The use of anoxic zones results in nitrate being reduced to gaseous nitrogen and hence the concentration of nitrate discharged in plant effluent is reduced.

(iii) The use of nitrate oxygen to satisfy BOD in anoxic zones results in a reduction in the overall oxygen requirements of the process and hence a saving in energy.

(iv) Problems associated with denitrification in final settlement tanks are minimised.

(v) Slime growths on the diffusers in the aerated zone downstream of the anoxic zone should be minimal.

Following the anoxic zone the aeration tank was divided into 3 separate zones by the use of baffles. Within these zones the layout of diffusers and the action of the control system was such that the dissolved-oxygen concentration was maintained at about 2 mg/1. This is considered adequate for the maximum rate of conversation of ammonia to nitrate by nitrifying bacteria.

The fourth aerated zone, occupying only a small fraction of the total tank volume would be used to maintain a higher concentration of dissolved oxygen (4 mg/1). It was recognised that a lower aeration efficiency would result in this zone but this would involve only low flow-rates of air since the design was such that treatment would normally be complete at the beginning of this zone. It has been shown that high concentrations of dissolved oxygen are desirable to stimulate the action of activated-sludge protozoa. Protozoa are instrumental in removing finely-divided suspended material of bacterial origin and their presence is necessary for the production of high-quality effluents. It would also provide extra aeration capacity in the event of an unexpectedly high demand for oxygen when the rate of supply in that zone could increase by about 60% corresponding to a decrease in dissolved oxygen to 0.5 mg/1.

The initial proposals for configuration of diffuser-domes in the non-nitrifying tank were similar to those for the nitrifying tank. However, in the former case air would be supplied to all five sections of the aeration tank with the first compartment used only for re-aeration of recycled sludge. Settled sewage would enter the aeration tank at the beginning of the second compartment as shown in Figure 4A(b).

In aeration tanks treating sewage at a high rate, as proposed in the non-nitrifying plant, it has often been found that microbial slimes grow on the diffusers near the tank inlet. This process causes clogging of the pores of the diffusers and the production of coarse bubbles. As a result there is a reduction in rate and efficiency of oxygen transfer and also an increase in back-pressure across the diffusers and hence energy is wasted and effluent quality deteriorates. It has been found in the laboratory that re-aeration of recycled sludge can prevent the sliming of the diffusers in the downstream when sewage is mixed with the re-aerated sludge.

It was considered that dissolved-oxygen concentrations of about 1 mg/l in the next 3 aerated compartments of the non-nitrifying plant would normally be adequate for carbonaceous oxidation. The final zone would be the same as in the nitrifying plant for the same reasons.

At present there is no reliable theoretical method for predicting the degree of aeration taper required in a plug-flow activated-sludge aeration tank with a given oxygen demand. The variation in oxygen demand along a tank is reflected by the change in respiration rate of sludge that occurs. The respiration rate is a function of the nature and biodegradability of the sewage treated and also depends on the length to width ratio of the aeration tank. A typical curve of respiration rate for a non-nitrifying sludge versus length along an aeration tank from inlet to outlet might look similar to that shown in Figure 5A(a). The respiration rate and hence the oxygen demand, decreases as treatment progresses, from an initially high value to a low level which, if treatment is complete, represents the endogenous respiration rate of the sludge.

The amount of air required in any particular zone of an aeration tank is therefore proportional to the relevant area beneath a respiration rate curve such as that shown in Figure 5A(a). Unfortunately it is impossible to derive accurately such curves before the aeration tank has been constructed. This presents no disadvantage where exact replicas of existing tanks are being added to a works to accommodate increases in flow but presents problems for a new works design or for the design of aeration tanks to operate under substantially different conditions from existing aeration tanks. In such cases the variation in demand shown in Table 1 can be used with reasonable confidence.

In a nitrifying plant the total oxygen requirement includes an amount of oxygen required for oxidation of ammonia. As the demand for oxygen for nitrification is normally exerted at a constant rate throughout the tank then the respiration rate curves for a nitrifying and a non-nitrifying aeration tank treating the same sewage might appear as shown in Figure 5A(b). The intensity of aeration required in a nitrifying plant is less

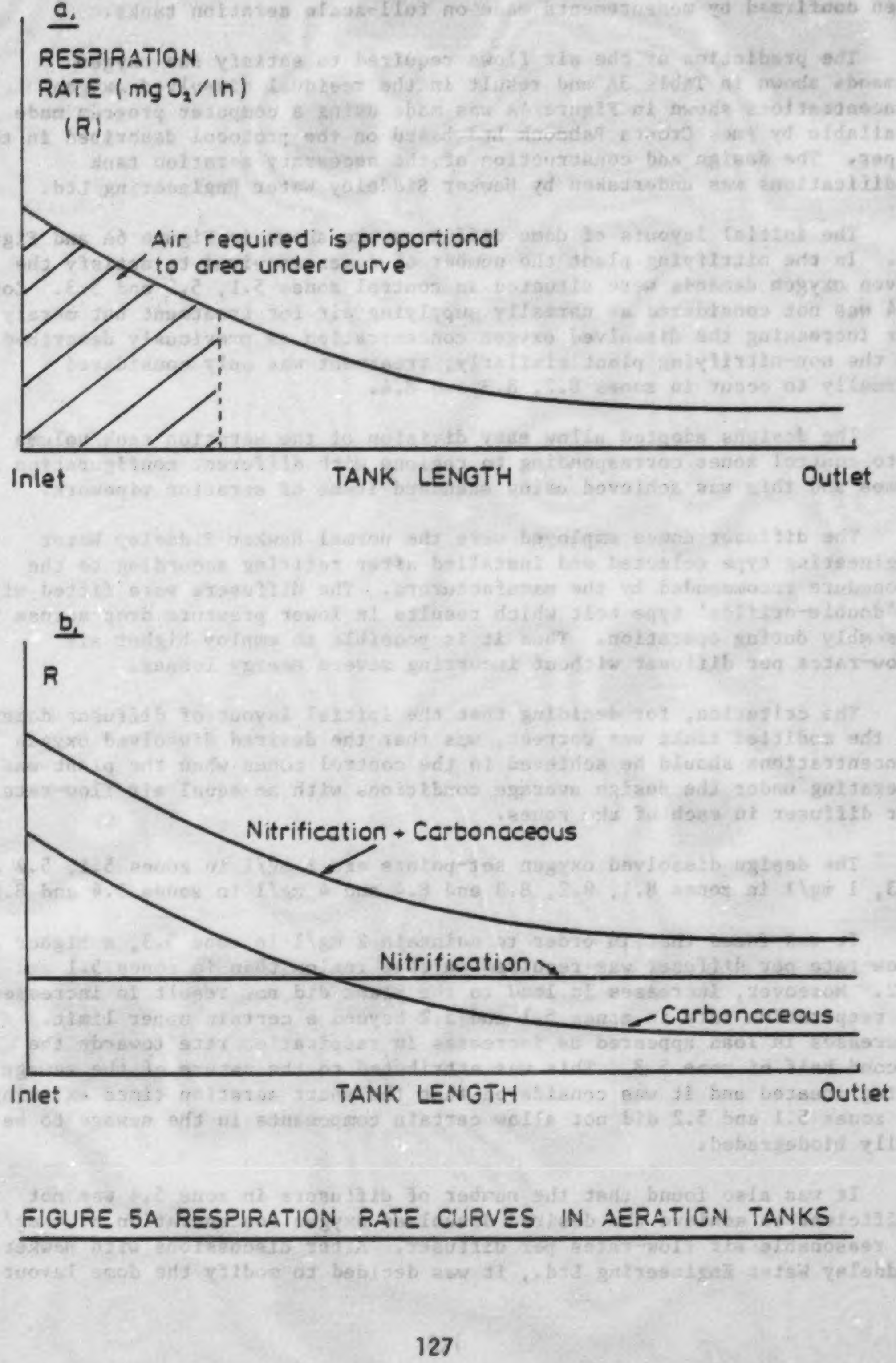


FIGURE 5A RESPIRATION RATE CURVES IN AERATION TANKS

extreme than that required in a non-nitrifying plant. This assumption has been confirmed by measurements made on full-scale aeration tanks.

The prediction of the air flows required to satisfy the oxygen demands shown in Table 3A and result in the residual dissolved oxygen concentrations shown in Figure 4A was made using a computer program made available by Ames Crosta Babcock Ltd based on the protocol described in this paper. The design and construction of the necessary aeration tank modifications was undertaken by Hawker Siddeley Water Engineering Ltd.

The initial layouts of dome diffusers are shown in Figure 6A and Figure 7A. In the nitrifying plant the number of domes required to satisfy the given oxygen demands were situated in control zones 5.1, 5.2 and 5.3. Zone 5.4 was not considered as normally supplying air for treatment but merely for increasing the dissolved oxygen concentration as previously described. In the non-nitrifying plant similarly, treatment was only considered normally to occur in zones 8.2, 8.3 and 8.4.

The designs adopted allow easy division of the aeration tank volume into control zones corresponding to regions with different configuration of domes and this was achieved using standard items of aeration pipework.

The diffuser domes employed were the normal Hawker Siddeley Water Engineering type selected and installed after refiring according to the procedure recommended by the manufacturers. The diffusers were fitted with a 'double-orifice' type bolt which results in lower pressure drop across the assembly during operation. Thus it is possible to employ higher air flow-rates per diffuser without incurring severe energy losses.

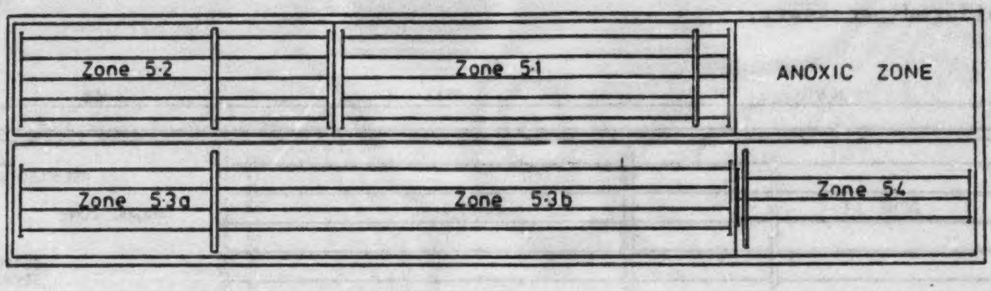
The criterion, for deciding that the initial layout of diffuser domes in the modified tanks was correct, was that the desired dissolved oxygen concentrations should be achieved in the control zones when the plant was operating under the design average conditions with an equal air flow-rate per diffuser in each of the zones.

The design dissolved oxygen set-points are 2 mg/l in zones 5.1, 5.2 and 5.3, 1 mg/l in zones 8.1, 8.2, 8.3 and 8.4 and 4 mg/l in zones 5.4 and 8.5.

It was found that in order to maintain 2 mg/l in zone 5.3, a higher air flow-rate per diffuser was required in that region than in zones 5.1 and 5.2. Moreover, increases in load to the plant did not result in increases in respiration rate in zones 5.1 and 5.2 beyond a certain upper limit. Increases in load appeared as increases in respiration rate towards the second half of zone 5.3. This was attributed to the nature of the sewage being treated and it was considered that the short aeration times existing in zones 5.1 and 5.2 did not allow certain components in the sewage to be fully biodegraded.

It was also found that the number of diffusers in zone 5.4 was not sufficient to achieve the desired dissolved oxygen concentration of 4 mg/l at reasonable air flow-rates per diffuser. After discussions with Hawker Siddeley Water Engineering Ltd., it was decided to modify the dome layout in

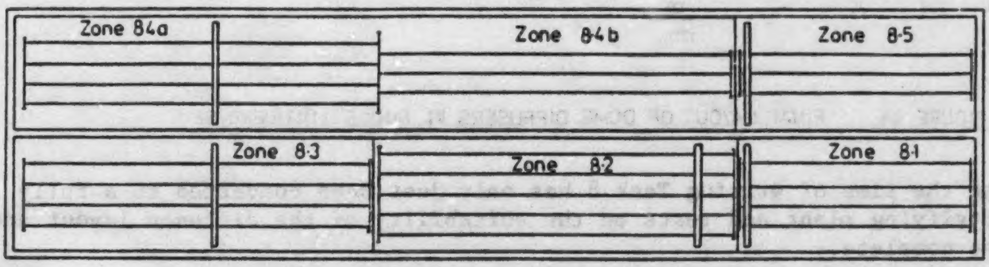
FIGURE 6A INITIAL LAYOUT OF DIFFUSERS IN AERATION TANK NO 5 (NITRIFYING)


Zone 5·2
Zone 5·1
ANOXIC ZONE
Zone 5·3a
Zone 5·3b
Zone 5·4

| ZONE | DOME LAYOUT | TOTAL NO. DOMES | % TOTAL |
|---|---|---|---|
| 5·1 | 5 Rows 88 per row | 440 | 36·8 |
| 5·2 | 5 - 64 - " | 320 | 26·8 |
| 5·3a | 4 - 64 - " | 256 | 21·3 |
| 5·3b | 4 - 45 - " | 180 | 15·1 |
| 5·4 | 3 - 25 - " | 75 | |
| | | 1196 | 100·0 |
| | | 1271 | |

| | Max. | Median | Min. |
|---|---|---|---|
| Oxygen demand (kg/d) | 4910 | 1925 | 1120 |
| Total air flow (l/s) | 1525 | 525 | 290 |
| Air flow/diffuser (l/min) | 77 | 26 | 15 |

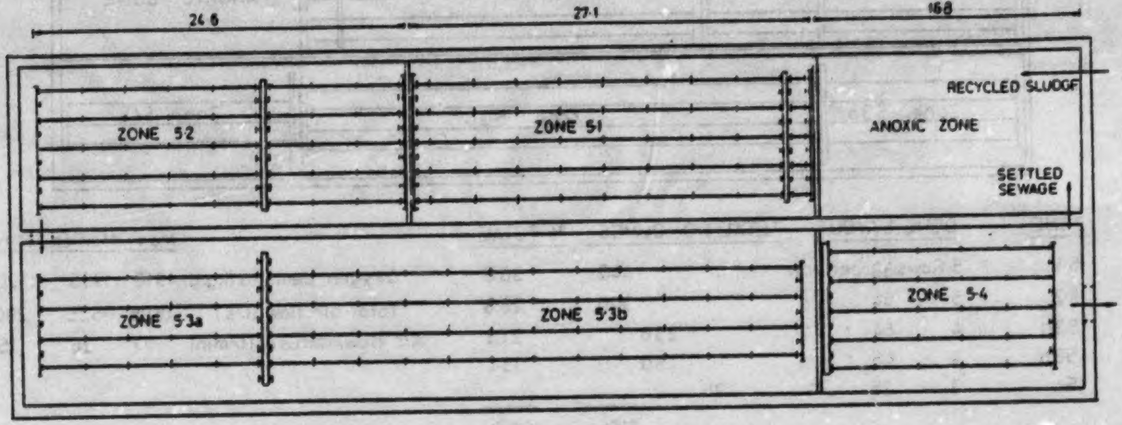
FIGURE 7A INITIAL LAYOUT OF DIFFUSERS IN AERATION TANK NO 8 (NON-NITRIFYING)


Zone 8·4a
Zone 8·4b
Zone 8·5
Zone 8·3
Zone 8·2
Zone 8·1

| ZONE | DOME LAYOUT | TOTAL NO. DOMES | %TOTAL |
|---|---|---|---|
| 8·1 | 4 Rows, 25 per row | 100 | |
| 8·2 | 5 " 88 " | 440 | 47·5 |
| 8·3 | 4 - 48 - | 192 | 20·7 |
| 8·4a | 4 - 40 - | 295 (160 | 17·3 |
| 8·4b | 3 - 45 · | (135 | 14·5 |
| 8·5 | 3 - 25 - | 75 | |
| | | 927 | 100·0 |
| | | 1102 | |

| | Max. | Median | Min. |
|---|---|---|---|
| Oxygen demand (kg/d) | 3690 | 1320 | 675 |
| Total air flow (l/s) | 1365 | 400 | 185 |
| Air flow/diffuser (l/min) | 88 | 26 | 12 |

zones 5.3.b and zone 5.4. Accordingly, Tank 5 was drained and additional domes were added to give the layout shown in Figure 8A. Subsequent operation of the plant has revealed that the layout of the diffusers is now very close to that required and hence no further modifications were considered necessary.



| ZONE | DOME LAYOUT | TOTAL NO. DOMES | % OF TOTAL |
|---|---|---|---|
| 5·1 | 5 ROWS 88 PER ROW | 440 | 35·1 |
| 5·2 | 5 - 64 - - | 320 | 25·6 |
| 5·3a | 4 - 64 - - | 256) )492 | 20·5 |
| 5·3b | 4 - 59 - - | 236) | 18·8 |
| | | 1252 | 100·0 |
| 5·4 | 5 - 25 - - | 125 | |
| | | 1377 | |

| | MAX. | MEDIAN | MIN. | |
|---|---|---|---|---|
| OXYGEN DEMAND | 4910 | 1975 | 1120 | kg/d |
| TOTAL AIR FLOW | 1525 | 525 | 290 | l/s |
| AIR FLOW/ DOME | 73 | 25 | 14 | l/min |

FIGURE 8A FINAL LAYOUT OF DOME DIFFUSERS IN TANK 5 (NITRIFYING)

At the time of writing Tank 8 has only just been converted to a fully non-nitrifying plant and tests on the suitability of the diffuser layout are not yet complete.

## A.5. PROCESS CONTROL, MONITORING, DATA LOGGING AND DATA PROCESSING SYSTEMS

### A.5.1. Introduction

The dissolved-oxygen control system and process monitoring arrangements associated with the project are more complex than those that would be required were the design philosophy to be replicated at another site. Thus the computer installed on site is only necessary to handle large quantities of on-line data and provide a degree of data processing which would be impractical manually.

A schematic diagram of the overall system is shown in Figure 9A. Process instrumentation provides information about the plant operating conditions to the computer via a telemetry system. The programmable logic controllers (PLC's) which are responsible for maintaining the dissolved oxygen at the required concentrations receive information from other instrumentation and take the necessary control action independently of the computer. The link from the PLC's to the computer is only necessary to provide the computer with data for processing purposes.

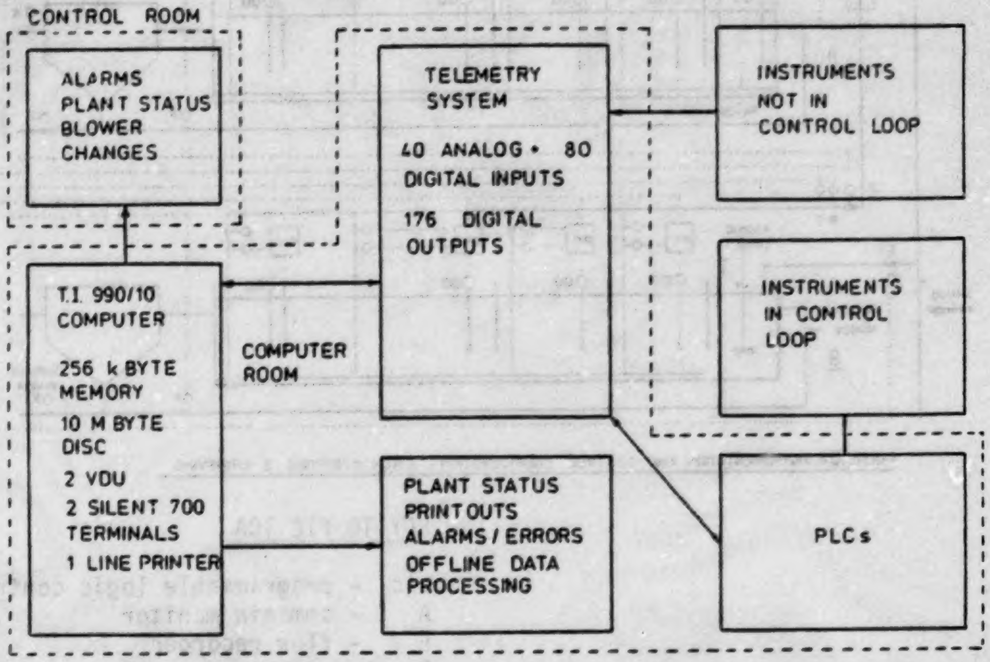

CONTROL ROOM
ALARMS
PLANT STATUS
BLOWER
CHANGES
TELEMETRY
SYSTEM
40 ANALOG • 80
DIGITAL INPUTS
176 DIGITAL
OUTPUTS
INSTRUMENTS
NOT IN
CONTROL LOOP
T.I. 990/10
COMPUTER
256 k BYTE
MEMORY
10 M BYTE
DISC
2 VDU
2 SILENT 700
TERMINALS
1 LINE PRINTER
COMPUTER
ROOM
INSTRUMENTS
IN CONTROL
LOOP
PLANT STATUS
PRINTOUTS
ALARMS / ERRORS
OFFLINE DATA
PROCESSING
PLCs

FIGURE 9A PROCESS CONTROL AND DATA PROCESSING BLOCK DIAGRAM

The computer processes the information received and provides various summaries, some of which are directed to the Rye Meads Control Room for the information of Thames Water Staff and some of which are for direct comparisons of plant performance.

The instrumentation and control configuration for the two modified plants is shown in Figure 10A. Each zone of the aeration tanks (as described previously) has a dissolved-oxygen probe, a separate, metered, controllable air supply and a PLC.

The air supply to the aeration tanks is provided by a set of six fixed-speed Rootes blowers, each of which is rated at about 1650 l/s (at NTP). The design of Rootes blowers is such that they can be considered as constant-volume machines (i.e. positive displacement). Thus increases in resistance to flow downstream of the blowers are compensated by increases in output pressure such that similar volumetric output is maintained.

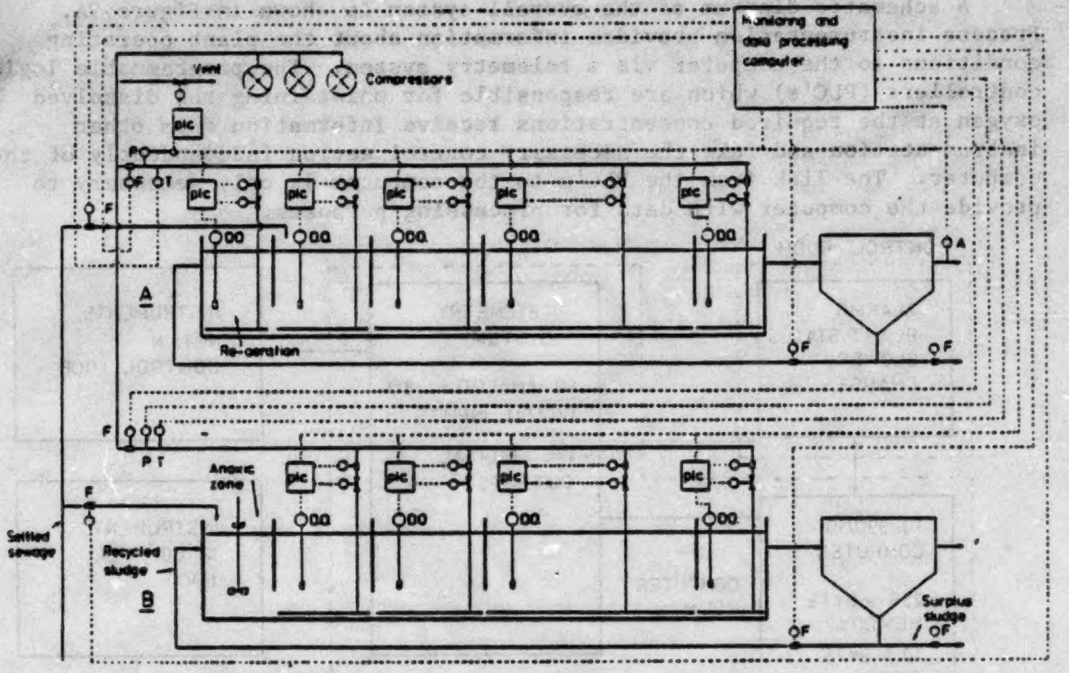

Monitoring and data processing computer
Vent
Compressors
plc
P
F
T
DO
A
Re-aeration
F
P T
Anoxic zone
Settled sewage
Recycled sludge
B
Surplus sludge
F

FIGURE 10A INSTRUMENTATION AND CONTROL CONFIGURATION - A.NON NITRIFYING B. NITRIFYING

KEY TO FIG 10A

pic - programmable logic controller
A - ammonia monitor
F - flow recorder
P - pressure recorder
T - temp. recorder
0.0. - dissolved oxygen recorder

**It was not considered justified for the purposes of this project to install a variable-speed compressor in the system. Hence the required variations in output are provided by the vent-valve shown in Figure 10A. With this arrangment the fixed volumetric output of the compressors is modulated by the vent-valve to provide a variable volume air supply to the aeration tanks.**

### A.5.2. Dissolved Oxygen Control System

The functions of the dissolved oxygen control system are:-

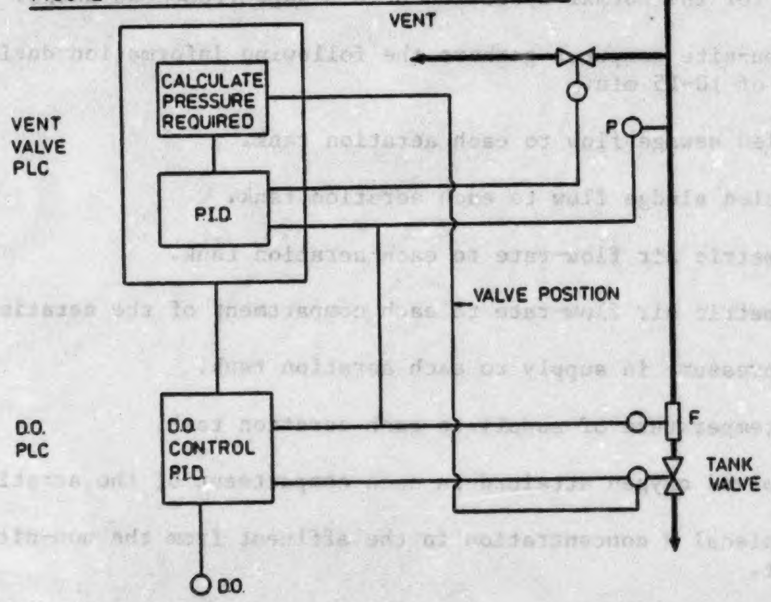
(i) to control the dissolved oxygen concentration in each compartment of each aeration tank to within ± 0.5 mg/1 of the design DO concentration; and

(ii) to achieve the required control such that the energy used for aeration is minimised irrespective of variations in demand for dissolved oxygen.

In addition the control system is also able to provide information for monitoring and data processing purposes.

The layout of the DO control PLC's is shown in Figure 10A and the detailed arrangement of the control loop which allows the vent valve arrangement to interact with the control system is shown in Figure 11A.

FIGURE 11A VENT VALVE AND DO CONTROL LOOPS



It is recognised that the number of DO control PLC's installed may be greater than the number required by a potential replicator but they were included to obtain evidence of the number actually needed to achieve satisfactory DO control.

Each control PLC compares the DO concentration measured in the appropriate aeration zone with the required set point. The errors in this comparison are then used to calculated desired valve positions using a PID control algorithm derived for this application by Hawker Siddeley Dynamics Engineering Ltd. This control loop cycles every 20-40 seconds.

The fixed volume characteristics of the blowers means that changes in tank valve positions do not result in the desired changes in air flowrates unless the system interacts with the vent-valve. The PLC which controls the vent-valve cycles every 5-10 seconds around another control loop and the vent valve is opened to the correct position such that a constant pressure in the air supply main results.

### A.5.3. Monitoring, Data Logging and Data Processing

In order that the objectives of the project are met it is necessary to collect data regarding the operation of the modified aeration tanks. An accurate economic comparison between the nitrifying plant and the non-nitrifying plant and also the unmodified control plant requires the installation of instrumentation and monitoring equipment which would not be necessary for the normal operation of a sewage treatment works.

The on-site computer gathers the following information during a cycle time of 10-15 min.

Settled sewage flow to each aeration tank.

Recycled sludge flow to each aeration tank.

Volumetric air flow-rate to each aeration tank.

Volumetric air flow-rate to each compartment of the aeration tank.

Air pressure in supply to each aeration tank.

Air temperature of supply to each aeration tank.

Dissolved oxygen attained in each compartment of the aeration tanks.

Ammoniacal N concentration in the effluent from the non-nitrifying plant.

Volumetric air flow-rate used for aeration of channels carrying mixed-liquor to sedimentation tanks.

Position of air vent-valve.

The data gathered by the computer are continuously displayed in the computer room. An example of this display is shown in Table 4A. A print-out of plant conditions similar to that shown in Table 4A is also produced every cycle on a terminal situated in the plant control room. This terminal is also used to signal major faults to the control room staff and to request changes in the number of compressors in use.

## Table 4A. Display of plant status

### Status of tanks 5 & 8 at 15.39, 23 April 1982

| | Tank 5 Nitrifying | | | | Tank 8 Non-nitrifying | | | | |
|---|---|---|---|---|---|---|---|---|---|
| Zone | 1 | 2 | 3 | 4 | 1 | 2 | 3 | 4 | 5 |
| DO Actual | 2.2 | 2.0 | 1.9 | 4.3 | 0.9 | 0.9 | 1.1 | 1.1 | 4.2 |
| Previous | 2.2 | 2.1 | 2.0 | 4.2 | 0.9 | 1.0 | 1.0 | 1.1 | 4.0 |
| Previous | 2.1 | 2.2 | 2.0 | 4.1 | 0.8 | 1.1 | 1.0 | 1.0 | 3.9 |
| Previous | 2.0 | 2.1 | 1.9 | 4.0 | 0.9 | 1.0 | 0.9 | 0.9 | 3.8 |
| Previous | 1.9 | 2.0 | 1.9 | 4.1 | 1.0 | 0.9 | 1.0 | 0.9 | 3.9 |
| Air flow 1/s | 183 | 133 | 205 | 52 | 41 | 182 | 80 | 123 | 31 |
| Previous | 180 | 145 | 212 | 57 | 45 | 187 | 82 | 131 | 35 |
| Previous | 190 | 165 | 215 | 53 | 40 | 195 | 78 | 128 | 28 |
| Previous | 185 | 140 | 223 | 55 | 43 | 180 | 75 | 125 | 25 |
| Previous | 186 | 130 | 217 | 48 | 45 | 183 | 77 | 121 | 30 |
| Air supply P | | | 4.2 | M.WG | | 4.2 | M.WG | | |
| Air supply flow | | | 575 | 1/s | | 460 | 1/s | | |
| Air supply T | | | 42.0 | DEGC | | 42.0 | DEGC | | |

ML Air flow rate 5 & 8 Total 65. 1/s

Sewage Flow 1/s Tk 5 90.2 88.5 84.2 87.1 Tk 8 121.0 118.6 122.3 125.9

RAS Flows 87. 1/s 120. 1/s

Air vent valve posn 45% Open Tk 8 Effluent ammonia concn 15. mg/1

No. compressors in use 1.

Analytical data pertaining to the operation of the modified aeration tanks are archived on a disc storage system by manual input. All the relevant data relating to the unmodified control plant are also archived manually. These data include:

(i) the results of analysis of composite samples of settled sewage and effluent for the nitrifying plant, non-nitrifying plant, and unmodified control plant. (The parameters measured are BOD, COD, TOC, total nitrogen, ammoniacal nitrogen, oxidised nitrogen, soluble phosphate, anionic surfactant and suspended solids);

(ii) the results of spot sample analysis to determine mixed liquor suspended solids, recycled sludge suspended solids, sludge settleability - as SSDI, sludge filtrability - as CST, and volatile matter in the mixed liquor suspended solids; and

(iii) the results of analysis for BOD and ammonia on two-hourly grab samples of settled sewage and for BOD, ammonia and oxidised nitrogen on two-hourly grab samples of each plant effluent.

A computer program has been written to allow derived parameters of interest for the comparison of the plants, to be calculated. The program involves interaction between manual input data and on-line data and permits calculation of the variability of parameters during the course of a day. For example, the variation in oxygen demand experienced by each modified plant during a day is calculated by combining the results of analysis of grab samples of settled sewage and effluent with corresponding values of settled sewage flow-rate.

A list of the derived parameters and their relevance to the project objectives is given below.

(i) % BOD Removal - This parameter allows comparison of the performance of each plant and is required to demonstrate that the modifications to the nitrifying plant do not result in any deterioration in effluent quality in comparison with the control plant.

(ii) Surplus sludge production - The increases in aeration efficiency resulting from the plant modifications may be offset if increases in surplus sludge production also result. The costs of sludge disposal are a major proportion of operating costs at sewage works and depend not only on the total amounts produced but also on their dewaterability. Hence the amounts of surplus activated sludge produced by the modified plants and the dewatering characteristics of the sludge are being closely monitored.

(iii) Oxygen demand - This parameter is a measure of the mass of oxygen that must be supplied via the flow of air to an aeration tank in order to satisfy the oxygen requirements as described by Equation 2. Oxygen demand varies with time according to the usual diurnal and seasonal variations in sewage flow-rate and composition. The oxygen demand satisfied by unit input of electrical energy used by the air blowers is the basic parameter for assessing the efficiency of the aeration systems.

(iv) The ratio of oxygen demand to oxygen supplied - This parameter provides a measure of the efficiency of the mass transfer of oxygen from the gas phase to the liquid phase in the aeration tank. The mass flow-rate of oxygen to the aeration tanks is calculated from on-line measurements of air flow-rate, temperature and pressure.

(v) Energy used for air compression - This parameter is a function of the mass flow-rate of air supplied to the aeration tanks and the pressure of the supply.

## A6. SUMMARY OF INITIAL RESULTS

Delays encountered during the commissioning of the control system have meant that to date few results have been obtained. At the present time Tank 8 has still not been completely converted to a non-nitrifying unit and the total amount of air produced by 1 compressor is being passed into the aeration tanks, i.e. the vent-valve and control system are not in operation. The results of approximately 1 month's operation of the plants are summarised in Table 5A. It can be calculated that the modified plants treat a far greater quantity of sewage (expressed in proportion to the relative aeration volumes) than the control plant; the percentage increase relative to the control plant was 84 for the nitrifying plant and 124 for the 'non-nitrifying' plant (which was partially nitrifying - Table 5A). Sludge settleability has also improved in the modified plants and there has been a slight reduction in the amount of surplus sludge produced.

The low aeration efficiencies reflect the lack of control of dissolved oxygen, however, under manual control of dissolved oxygen it has been possible to achieve an aeration efficiency of about 2 kg $O_2$/kWh in Tank 5.

Table 5A. Summary of average results obtained in modified plants - March 1982

| | Nitrifying<br>Tank 5 | 'Non-nitrifying'[1]<br>Tank 8 | Control plant<br>Tank 9 |
|---|---|---|---|
| **Settled sewage** | | | |
| BOD (mg/l) | 113 | 113 | 113 |
| COD (mg/l) | 300 | 300 | 300 |
| Ammoniacal-N (mg/l) | 32 | 32 | 32 |
| Soluble phosphate (mg P/l) | 7.4 | 7.4 | 7.4 |
| Anionic surfactant (mg/l) | 10.5 | 10.5 | 10.5 |

**Table 5A continued**

**Operating conditions**

| | | | |
|---|---|---|---|
| Volume of aeration tank ($m^3$) | 1900 | 1900 | 3800 |
| Settled sewage flowrate (l/s) | 88 | 106 | 96 |
| Recycled sludge flowrate (l/s) | 95 | 93 | 125 |
| MLSS (mg/l) | 5050 | 3790 | 4580 |
| Air flowrate (l/s) | 870 | 780 | 1130 |
| Surplus sludge flowrate (l/s) | 1.1 | 2.3 | 1.7 |
| Temperature of aerated liquid (°C) | 12.5 | 12.5 | 12.5 |
| **Effluent quality** | | | |
| BOD (mg/l) | 4.5 | 6.1 | 3.9 |
| COD (mg/l) | 38 | 41 | 30 |
| Ammoniacal-N (mg/l) | 0.6 | 5.7 | <0.2 |
| Oxidised nitrogen (mg/l) | 18.8 | 21.3[1] | 21.6 |
| Soluble phosphate (mg/l) | 8.7 | 8.6 | 9.2 |
| Detergent (mg/l) | 0.3 | 0.3 | <0.2 |
| **Derived data** | | | |
| SSDI (at 3.5 g/l) | 1.75 | 1.64 | 1.23 |
| Volumetric loading (l/s per 1000 $m^3$) | 46 | 56 | 25 |
| Sludge loading ($d^{-1}$) | 0.09 | 0.14 | 0.05 |
| Sludge age (d) | 9.1 | 4.6 | 13.1 |
| Sludge production (g/g BOD applied) | 1.22[2] | 1.55[2] | 1.53[2] |

| | Nitrifying Tank 5 | 'Non-nitrifying'[1] Tank 8 | Control plant Tank 9 |
|---|---|---|---|
| Hydraulic residence time (h) | 2.9 | 2.6 | 4.7 |
| Oxygen demand satisfied (kg/d) | 1840 | 2000 | 2400 |
| Theoretical power consumption (kWh/d) | 1240 | 1125 | 1885 |
| Aeration efficiency (kg $O_2$/kWh) | 1.5[3] | 1.8[3] | 1.3 |

1. Partial nitrification occurring.
2. These values are probably too high because the period for data collection was too short.
3. The results are low because the system to control flow-rate of air and maintain low dissolved-oxygen concentrations was not in operation.

# EXPERIENCES WITH DIFFERENT AERATION SYSTEMS IN GERMANY

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## ABSTRACT

Results from measurements under process conditions of the oxygen transfer and the aeration efficiency, in conjunction with experiences with different aeration systems in use in Germany have been examined. It is found that the aeration efficiency of well-designed diffused air and surface aeration systems is of the order of 1.5 to 2.0 kg $O_2$/kWh, with the exception of diffused air in spiral flow tanks, which has a considerably lower aeration efficiency.

For the selection of aeration systems numerous factors have to be considered. Since in future denitrification has to be considered it is shown that circulation tanks with surface aerators are preferable.

## MEASUREMENT OF THE OXYGEN TRANSFER UNDER PROCESS CONDITIONS

In order to measure the oxygen transfer rate in tap water or mixed liquor, the D.O. has to be unbalanced. In tap water the D.O. is usually brought to zero by the addition of sodium sulfite. Similary D.O. of the mixed liquor is reduced to zero by switching off the aeration. After restarting aeration in both clean water and mixed liquor similar reaeration curves are obtained. They can be evaluated by use of the same procedure.

An interruption of the aeration in circulation tanks with surface aerators is illogical since it takes several minutes to build up the circulation velocity after aeration has been restarted. Due to the higher differential velocity between water and aerator during this period the oxygen transfer is higher. Another problem associated with the reaeration of mixed liquor is that at high oxygen uptake rates the D.O. may only increase by 2-3 mg/l thus producing unreliable results. In order to overcome these problems the reaeration test was inverted: without stopping aeration hydrogen peroxide was added to the aeration tank, which immediately decays to water and oxygen thus raising the D.O. as desired by approximately 10 to 15 mg/l. By continued aeration the excess oxygen is stripped off and a desorption curve is obtained which is homologous to the reaeration curve (Figure 1). If the liquid film coefficient $k_L$ has the same value for oxygen absorption and desorption, respectively, the slopes of the absorption and desorption curves at the same value of the apparent D.O. saturation deficit ($c_s$*-c) and D.O. supersaturation (c-$c_s$*) must then have same value. The desorption tests therefore, can be evaluated like reaeration tests. The theory behind this is described in (1). For the evaluation we would like to thank Brown and Fisette (2) for the use of their computer program. The oxygen transfer rate under process conditions is calculated without considering the ß-factor. We use the clean water standard (book) saturation value.

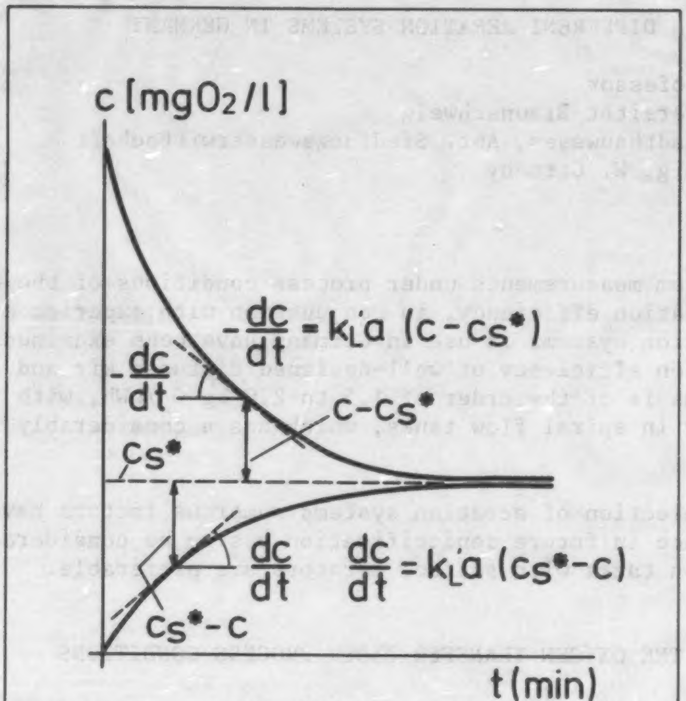
c [mgO2/l]
$-\frac{dc}{dt} = k_L'a\,(c - c_S^*)$
$-\frac{dc}{dt}$
c-cs*
cs*
$\frac{dc}{dt}$
$\frac{dc}{dt} = k_L'a\,(c_S^* - c)$
cs*-c
t (min)

Figure 1. Oxygen absorption (lower curve) and oxygen desorption (upper curve)

**The assumptions of testing with peroxide have been verified by performing a series of several tests with reaeration and desorption of D.O. in laboratory and full scale tanks.**

**Procedures for testing full scale tanks were developed once the theory was worked out. After first considering dilution of peroxide with tap water it was finally decided to pump technical grade 35% hydrogen peroxide directly into the aeration tank.**

**An even distribution of the peroxide in the whole aeration tank is essential for the reliability of the results. In tanks with cone aerators peroxide was dosed at each cone in mid-depth between the cone and the wall. In spiral flow tanks with diffused air the points of influx should be separated by a distance of about the tank width. The procedures for the other tank types shall not be stressed here.**

**In order to check the reproducability of the method, at the same plants with the same airflow the tests were repeated on different days. Values of SOTE obtained at one plant, for example, are: 8.8; 7.3; 7.3; 9.7; 9.1 and 9.1 g/m³·m. The average is 8.23 and the standard deviation 8.1%. The smallest value was 11% lower and the highest was 18% higher than the average. It has**

to be stressed that these measurements were performed in an extended aeration plant under full operating conditions without interrupting sewage and excess sludge flow.

When the sewage and return sludge flow was interrupted at another plant on two different days the measurements for one airflow yielded 7.7; 7.9 and 7.7 $g/m^3 \cdot m$ and for another airflow 9.3; 9.0 and 9.7 $g/m^3 \cdot m$. The results were less scattering than that from measurements under full operating conditions.

## REQUIRED OXYGEN TRANSFER

The oxygen uptake rate depends mainly on the loading conditions and the degree of nitrification. The loading rate to be selected is a function of the required effluent quality. In Germany the discharge requirements are laid down in a federal regulation: with respect to the biological process only BOD$_5$ is relevant: the BOD$_5$ of 24 hour composite samples should be below 20 mg/l 80% of the time. This means that in general the effluent BOD has to be in the range of 10 to 15 mg/l. In order to advise consultants and planners the German Water Pollution Control Federation (ATV) has issued guidelines for the design of activated sludge plants (3). Recommended loading rates and required oxygen transfer rates are collected in Table 1. The hourly $\alpha$SOTR is the value of the peak hour. Because most plants are not fully loaded when they start operation and because of the daily and weekly oscillation of the oxygen uptake rate it is recommended to be able to vary $\alpha$SOTR by about 5:1.

Table 1. Design proposal of ATV (Ref. 3)

| | | Extend Aeration | Nitrification | |
|---|---|---|---|---|
| | | | with | without |
| F/M | kg BOD/kg MLSS | 0.05 | 0.15 | 0.30 |
| Vol. load | kg BOD/m³·d | 0.25 | 0.5 | 1.0 |
| $\alpha$SOTR | kg O$_2$/m³·d | 0.625 | 1.25 | 1.5 |
| $\alpha$SOTR | g O$_2$/m³·h | 26 | 52 | 63 |
| MIN $\alpha$SOTR | g O$_2$/m³·h | 6 | 12 | 15 |

## AERATION EQUIPMENT AND TANK CONFIGURATIONS

### Diffused air (fine bubble)

In Germany diffused air aeration is employed both in large and small plants. The diffusers in general are located along one wall in order to ensure a spiral flow within the tank. The specific oxygen transfer rates as recommended in the ATV Handbook (4) are shown in Table 2. The author agrees with the tap water data but for process conditions in two different plants a value of about 5.5 g $O_2/(m^3$ air$)\cdot$(m diffuser depth) was found. Assuming a gross power requirement of about 5 Watthours/m³ air ·(m diffuser depth) the aeration efficiency becomes 1.1 kg $O_2$/kWh (1.8 lb $O_2$/hp-hr).

Table 2. Recommended oxygen transfer of fine bubble aeration in spiral flow tanks (Ref. 4)

| | Good conditions | | Fair conditions | |
|---|---|---|---|---|
| | SOTR[a] | AE[b] | SOTE[a] | AE[b] |
| Clean water | 12 (1.3) | 2.2 (3.6) | 10 (1.1) | 1.7 (2.8) |
| Process conditions | 10 (1.1) | 1.8 (3.2) | 8 (0.9) | 1.3 (2.1) |

[a] g $O_2/m^3$ air ·(m diff.depth) x 0.1089 = (% $O_2$/ft diff.depth)

[b] kg $O_2$/kWh x 1.643 = lb $O_2$/hp-hr

The disadvantage of diffused air in spiral flow tanks is well known: due to the similar direction of the water movement and the rising air bubbles the contact time of the bubbles is much lower than if they were rising in quiescent water.

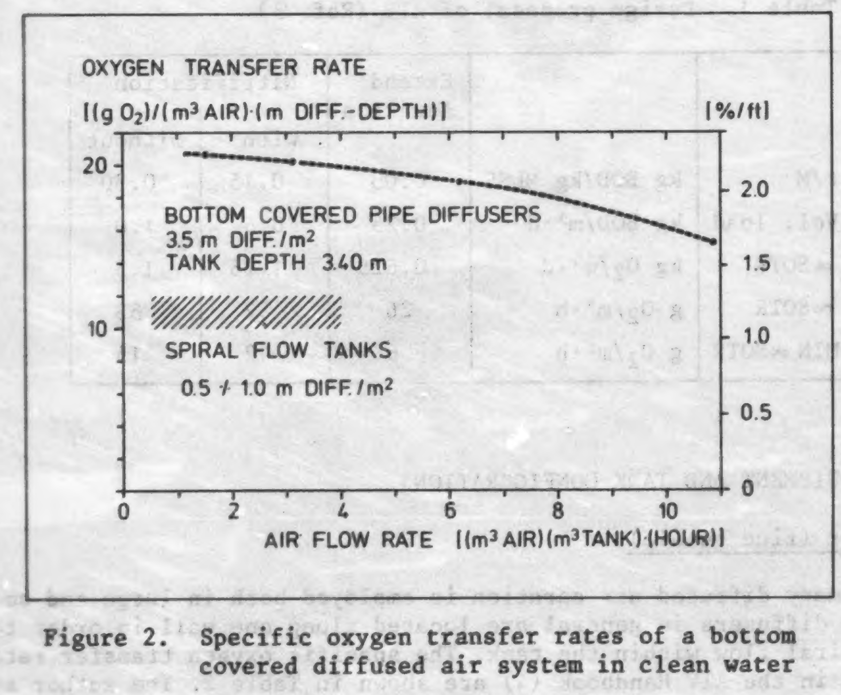

Figure 2. Specific oxygen transfer rates of a bottom covered diffused air system in clean water

Much better results therefore, are obtained when the diffusers are evenly distributed on the tank bottom (5). In Fig. 2 results from clean water measurements in a tank covered with tubular diffusers are compared with the generally accepted values for spiral flow tanks (6). The efficiency of the system tested was about 70 to 100% higher than that of the conventional spiral flow arrangement.

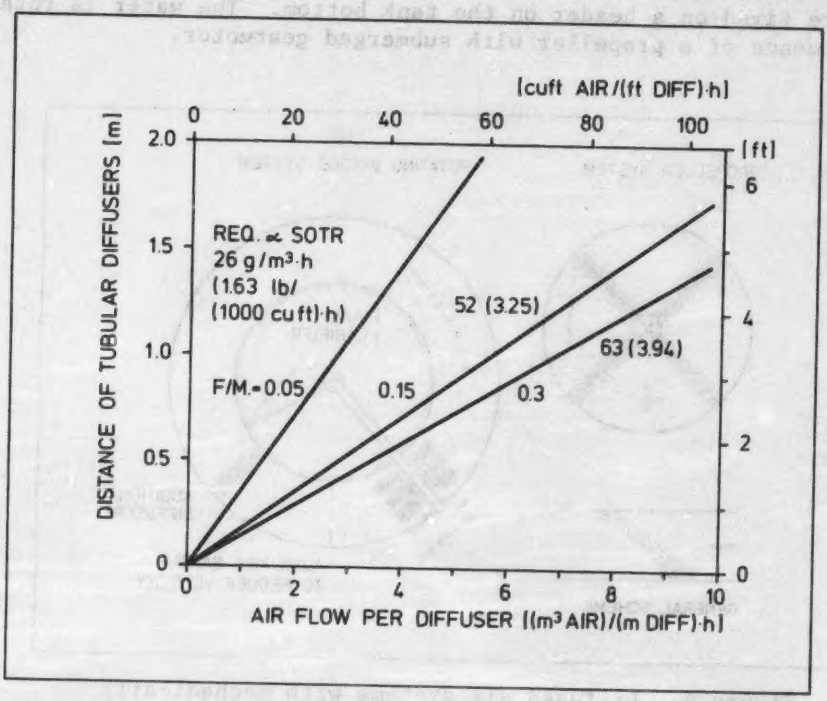

Figure 3. Diffuser spacing and air flow per diffuser for an assumed specific oxygen transfer rate of 10 $g/m^3 \cdot m$

In order to demonstrate the problems associated with an even diffuser distribution in Fig. 3 the dependence of diffuser spacing, air flow per diffuser and required oxygen transfer rate is shown. It was assumed that the oxygen transfer rate under process conditions is about 10 g $O_2/m^3$ air·m (~1%/ft,$\alpha$~0.5); values for the required oxygen transfer rate were taken from Table 1. As a design air flow per meter of tubular diffusers 8 to 10 $m^3$/h (80-100 cu ft/ft·hr) are used generally, which would lead to a distance between diffusers of about 1.5 m (5 ft) for plants with F/M = 0.3-0.15. Such an arrangement will create a flow pattern which might be closer to spiral flow than to a bottom cover type; furthermore, sludge deposits between the diffusers may not be avoided. If, on the other hand, a spacing of 0.5 m (~1.5 ft) is utilised the air flow per diffuser reduces to 3.0 $m^3/m \cdot h$ (30 cu ft/ft/hr) at F/M=0.3-0.15, or half the value for extended aeration. Subsequently the investment costs for diffusers and piping would be fourfold or more. Dome type diffusers may be more useful.

The investment costs however, will be of the same order of magnitude.

The problem was solved by Pasveer and Sweeris (7), who split aeration and mixing. At flow velocities between 0.3 and 0.5 m/s (1.0 and 1.5 ft/sec) they found in a one meter (3 ft) deep tank oxygen transfer efficiencies of about 30 $g/m^3 \cdot m$ (3%/ft). The two systems based on this principle which are manufactured in Germany are shown in Fig. 4. In the tank on the left side the diffusers are fixed on a header on the tank bottom. The water is rotated in the tank by means of a propeller with submerged gearmotor.

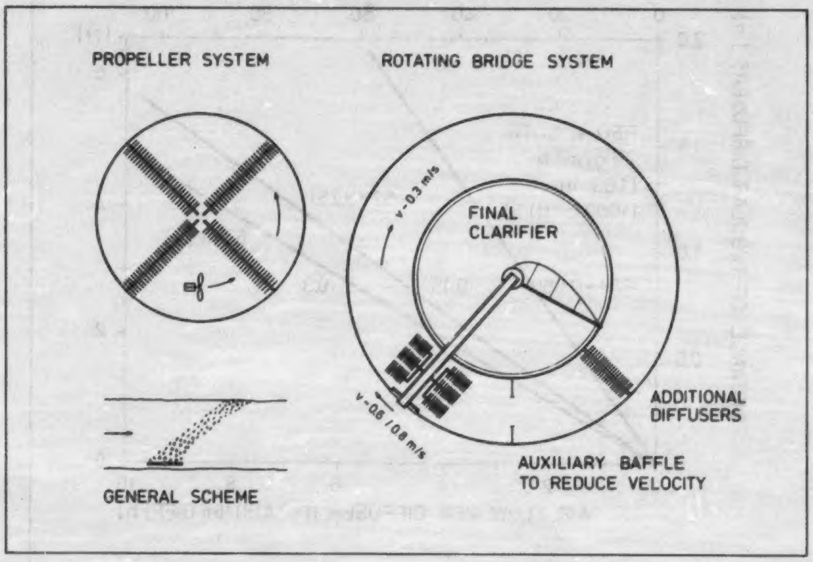

Figure 4. Diffused air systems with mechanically forced water flow

The other system was used only for extended aeration in past. The aeration tank is constructed around the final clarifier. The diffusers are fixed at the bridge which rotates with a peripheral velocity of 0.6 to 0.8 m/s (2-2.5 ft/sec). Due to the rising bubbles the water in the circular tank rotates at about 0.3 m/s depending on the air flow, thus leaving a differential velocity of 0.3 to 0.5 m/s. In larger tanks a baffle has to be installed in order to reduce the water velocity. In plants with higher oxygen uptake rates additional bottom-fixed diffusers have to be installed which reduce the water velocity; a baffle in this case is not needed.

In both plant types we made oxygen transfer measurements under process conditions (Table 3). The oxygen transfer efficiency of the propeller type plant M was excellent but the aeration efficiency was not as good as one may have expected because the propeller draw 2.5 kW and the blowers were overdesigned with respect to pressure. In plant H in addition to clogged diffusers the propellers draw 7 and 5 kW, therefore the efficiency is as low as in spiral flow systems. It has to be mentioned that the manufacturer of this system has improved the propeller. They claim that with the new propeller the power requirement for circulation is only 1 W/m³ (0.4 hp/1000 cu ft).

Table 3. Results from measurements under process conditions for four plants with diffused aeration

| | | Propeller system | | | | Rotating bridge | | | |
|---|---|---|---|---|---|---|---|---|---|
| | | M | | H[a] | | N | | | B |
| Tank diam. | (m) | 15.5 | | 17.9 | | 26/38 | | | 22/38 |
| Tank vol. | $(m^3)$ | 632 | | 860 | | 1735 | | | 2300 |
| Diff. length | (m) | 52 | | 66 | | 72 + 120[b] | | | 90 |
| Diff. depth | (m) | 3.05 | | 3.2 | | 2.8 | | | 2.8 |
| Air fl. rate[c] | $(m^3/m^3 \cdot h)$ | 0.8 | 0.4 | 1.4 | 0.6 | 1.1 | 0.7 | 0.4 | 0.4 |
| per diff. | $(m^3/m \cdot h)$ | 9.6 | 4.8 | 18 | 7.5 | 9.5 | 6.4 | 3.2 | 10 |
| $\alpha$SOTR | $(g/m^3 \cdot h)$ | 34 | 21 | 28 | 14 | 23 | 18 | 14 | 9.3 |
| $\alpha$SOTE | $(g/m^3 \cdot m)$ | 14.2 | 17.8 | 6.3 | 7.7 | 7.8 | 9.1 | 13.9 | 8.2 |
| P[d] | (kW) | 13.9 | 8.8 | 27.2 | 12.5 | 27.5 | 19.3 | 10.7 | 17.1 |
| AE | (kg/kWh) | 1.56 | 1.54 | 0.95 | 1.0 | 1.44 | 1.62 | 2.26 | 1.25 |

[a] clogged diffusers

[b] Fixed on bridge + fixed on the bottom

[c] Air discharge from blower nameplate

[d] Wire energy uptake of blower plus propeller or bridge

The rotating bridge of both plants N and B drew approximately 2.0 to 2.5 kW, which is close the value of 1 W/m³ mentioned previously. The oxygen transfer efficiency however, is much lower than that of the propeller system (14.2 - 17.8 $g/m^3 \cdot m$ as compared with 7.8 - 13.9 $g/m^3 \cdot m$). Only when the air flow rate is far below the design capacity, the aeration efficiency is good. The reason for the low efficiency at high air flow rates is the high air load at one point (the bridge). In smaller plants with tank volumes below 500 m³ the aeration efficiency will be better because the air load at the bridge is much lower.

Another invention comes close to a perforated tank bottom; steel boxes having a perforated rubber sheet on the upper side about 0.25 cm thick. Recommended air flow rates are 0 to 3 $m^3/m^2 \cdot h$ (0 to 10 cu ft/sq ft/hr). In tap water tests the results were very promising; the transfer efficiency was about 30 $g/m^3 \cdot m$ (3%/ft). Because of the small bubbles however, the $\alpha$-factor was close to 0.3. The efficiency under process conditions therefore, was of the same order to magnitude as other diffused fine bubble systems, excepting the spiral flow system which may run at an efficiency which is 50% or more lower.

The so called "tower" aeration tanks which have been developed by the chemical industry in which the air flow is low due to the depth and therefore, the treatment of exhausted air becomes cheaper, will not be stressed here.

## Vertical shaft surface aerators

After the first cone aerator was imported from England in 1962, another cone-aerator has been thrown in the market every year. By the end of the sixties Germany had become the "cone-country".

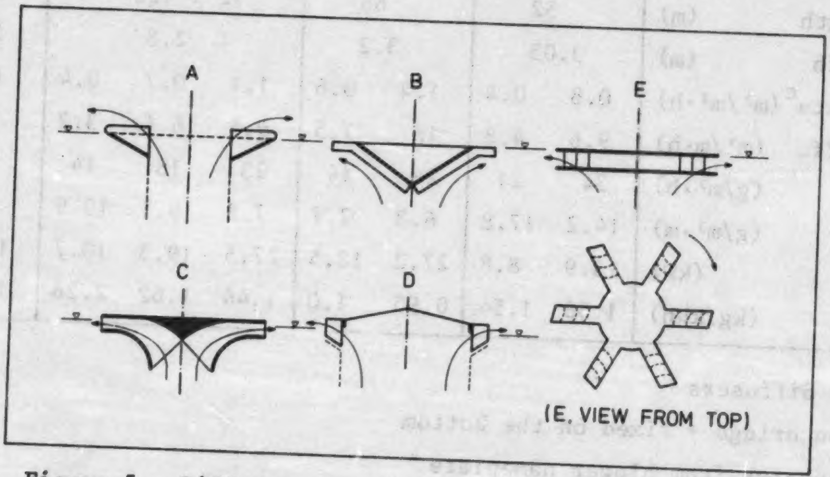

Figure 5. Different types of vertical shaft surface aerators (schematic)

Five typical cones are shown schematically in Fig. 5. Cone A (which is in principle the old Simplex-aerator) and cone D may or may not be used with a draft-tube. It is interesting to notice that one manufacturer changed the design of his aerator from type C to type A. The headlosses within aerators of type C are considerable due to the high velocities. Cones B, E and D (without a draft tube) may create higher turbulence in the aeration tank because their blades dip right into the water.

Oxygen transfer and aeration efficiency depend on numerous factors. Measurements in clean water and under process conditions indicate that the value of $\alpha$ is close to unity. ATV (4) recommends the aeration efficiency be 1.8 to 2.2 kg/kWh for good and 1.3 to 1.8 kg/kWh for fair conditions (3.0 - 3.6 and 2.1 - 3.0 lb/hp-hr).

That manufacturers tend to overestimate the capability of their products can be demonstrated by results from tests of three aerators. The City of Hamburg planned to install 36 cone aerators in a treatment plant currently under construction. The aerators were to be equipped with two-speed motors. The maximum oxygen transfer required was 220 kg/h (485 lb/hr), at high speed, at low speed it should be half of that. Because of the high investment costs the three manufacturers (out of six) which tendered the most reasonable bids were invited to perform tests in advance, all in the same tank. Two aerators were modified B-types and one was a C-type. The mechanical data are as follows:

B1: diam. 3.25 m; speed 33.9 and 25.4 per min
B2: diam. 3.25 m; speed 39.7 and 29.7 per min
C : diam. 2.86 m; speed 41.06 and 30.85 per min

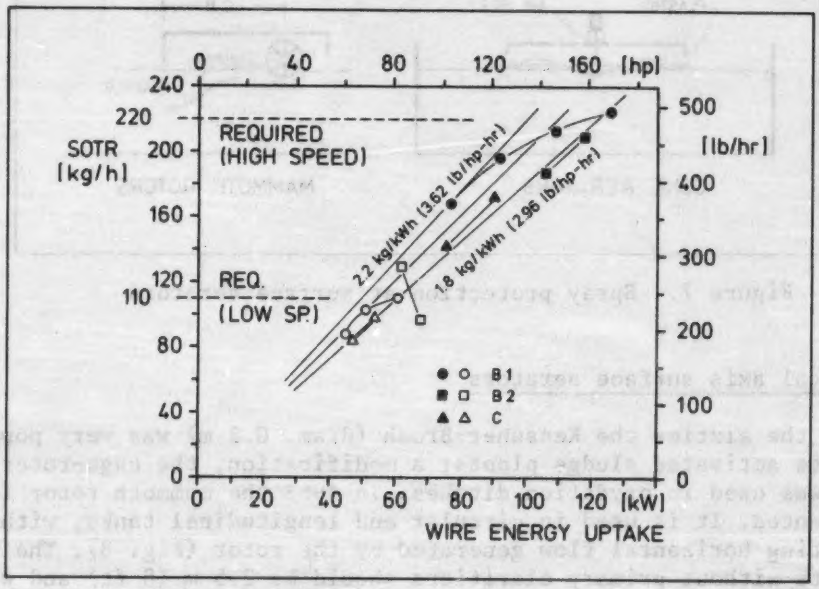

Figure 6. Results from clean water measurements with three vertical shaft surface aerators

The results are shown on Fig. 6. The tank volume was about 1000 m³. In principle all aerators failed. The one which just reached the 220 kg/h had its highest efficiency when it was less deeply immersed. In the authors opinion an appropriate diameter for all three cones would have been 3.5 m. The data however, indicate that proper vertical shaft aerators have a reasonable aeration efficiency of 1.8 kg/kWh (3.0 lb/hp-hr) and more.

The oxygen transfer of cone aerators may be controlled by variation of the depth of immersion and/or the speed. Variations of the water level do create shock loads in the final clarifiers. It is therefore, favourable to adjust the water level from time to time only and to control the daily fluctuations by speed variation using two or three speed motors, or as in the main Vienna treatment plant, by using fully variable-speed motors.

Because of the success of circulation tanks with cage and mammoth rotors, in the Netherlands the Carrousel tank with cone aerators was developed. Fig. 8 shows both a mammoth rotor and a Carrousel tank as well.

Other general problems associated with surface aerators are spray and noise emissions. Today a splash area should be covered (Fig. 7). The City of Hamburg for example, covers its tanks with cone aerators completely.

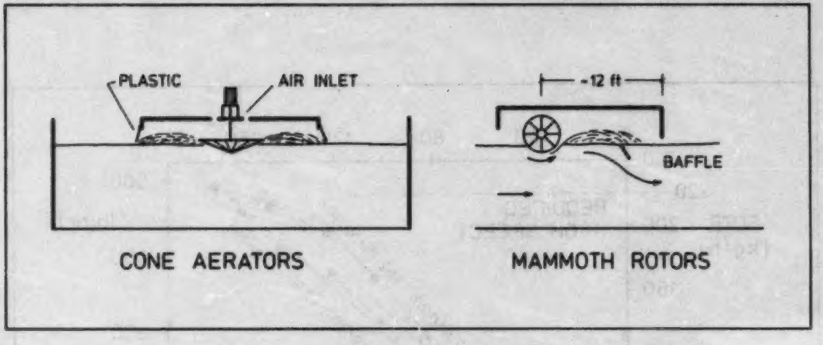

Figure 7. Spray protection at surface aerators

## Horizontal axis surface aerators

In the sixties the Kessener-Brush (diam. 0.5 m) was very popular for high rate activated sludge plants; a modification, the cage-rotor (diam. 0.7 m) was used in oxydation ditches. In 1963 the mammoth rotor (diam. 1.0 m) was invented. It is used in circular and longitudinal tanks, with the circulating horizontal flow generated by the rotor (Fig. 8). The tank depth of plants without primary clarifiers should be 2.5 m (8 ft) and with primary sedimentation the depth may be 3 to 4 m (10-13 ft). The maximum depth of immersion is 0.3 m (1 ft). In order to achieve a better distribution of the oxygen-enriched water over the tank depth downstream, a baffle is needed (Fig. 7). This baffle reduces the water velocity, which is necessary in tanks with more than two rotors. Because there may still remain a certain gradient of the oxygen concentration over the depth it is preferable to keep the separation between rotors at 30 m (100 ft) or more. Measurements in clean water and under process conditions indicated that $\alpha$ is close to unity. The oxygen transfer per meter of rotor per hour is 7 to 9 kg/m·h (4.7 - 6.0 lb/ft/hr) and the efficiency 1.8 to 2.0 kg/kWh (3.0 - 3.3 lb/hp-hr).

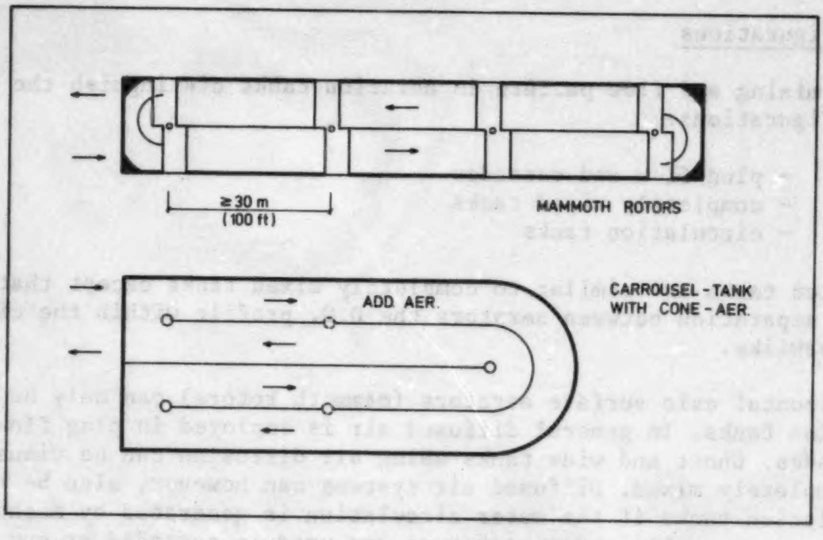

Figure 8. Circulation tanks (schematic)

We measured the oxygen transfer in one tank of the Braunschweig treatment plant under process conditions. The tank volume is 8700 m³ (307,000 cu ft) and it is equipped with 7 double rotors (2 x 7.5 m or 2 x 24.6 ft) each driven by a 75 kW motor (100 hp). A baffle is installed downstream of each rotor. The results are shown in Table 4, they clearly indicate that with an increasing number of rotors in operation the oxygen transfer and the efficiency decrease. That this is due mainly to the decreasing velocity between the rotor and the water is indicated by the decreasing power uptake per motor. Some sort adjustable baffle has to be developed by which the water velocity can be controlled. The oxygen transfer is most easily controlled by switching rotors on and off.

Table 4. Results from measurements with mammoth rotors under process conditions

| Rotors in operation | 7 | 5 | 4 |
|---|---|---|---|
| Total length of rotors (m) | 119 | 85 | 68 |
| $\alpha$SOTR (kg/m·h) | 6.45 | 7.00 | 8.85 |
| $\alpha$AE (kg/kWh) | 1.63 | 1.71 | 1.84 |
| Power upt. per motor (kW) | 59.6 | 61.5 | 72.3 |

### Tank configurations

The mixing and flow pattern in aeration tanks distinguish the following tank configurations:

- plug flow and cascades
- completely mixed tanks
- circulation tanks

Circulation tanks are similar to completely mixed tanks except that due to the wide separation between aerators the D.O. profile within the circulation tank is sawlike.

Horizontal axis surface aerators (mammoth rotors) can only be used in circulation tanks. In general diffused air is employed in plug flow tanks and cascades. Short and wide tanks using air diffusion can be visualized as being completely mixed. Diffused air systems can however, also be applied in circulation tanks if the water circulation is generated by mechanical means, e.g. a propeller. Cone aerators are used in cascaded or completely mixed tanks and in circulation tanks (refered to as Carrousel).

## SELECTION OF AERATION SYSTEMS

### General considerations

Criteria for selecting aeration systems are adequately outlined by v.d. Emde (8). A step-by-step procedure of some aspects of the decision making will be outlined in this chapter.

### Diffused air or surface aerators?

The recommended tank depth for surface aerators is in the range of 2.4-4.6 m (8-15 ft). In tank deeper than 4.6 m (15 ft) diffused air or other air aeration systems must be installed.

Diffused air systems seem to be preferable because of the slight heating effect of the compressed air in areas with a cold climate especially for small plants with long aeration times. In larger plants however, surface aerators can be covered (as shown in Fig. 7) in order to prevent ice from growing on the aerator or the structure. If nitrification is required even during strong winter periods diffused air systems may be superior. Consequently the headers should be submersed in order to lose as little heat as possible.

Certain wastewater constituents may cause clogging of diffusers. It has to be considered that small filaments tend to spin together even after the best screening and form rags which may wind around the diffusers if no primary sedimentation is employed. Only at type C cone aerators (Fig. 5) clogging by rags was observed.

If the exhaust air is to be treated deep tanks with diffused or other air systems are preferable. An elegant method of air treatment is being installed by the City of Frankfurt where a two stage activated sludge plant (both stages with diffused air systems, tank depth 20 ft) is under construction. The first stage aeration tank will be covered and the resulting spent air and some fresh air will be used as the air supply for the blowers of the second stage.

The variation of the oxygen supply by switching on and off blowers or motors is possible with diffusion in any type of tank and with surface aerators in circulation tanks. If cone aerators are installed in tanks in series each aerator has to be independantly adjustable (immersion and/or speed) for economical operation. A similar problem is the distribution of the air along aeration tanks.

Noise and spray problems of surface aerators can be solved by an appropriate cover of the splashing area and if necessary by a hood over the motor.

Smaller plants can be aerated by some means of compressed air only, because surface aerators for appropriately small oxygen transfer capacities are not available. In order to reduce operational problems big bubble aeration is preferable for very small plants.

Cost comparisons under German conditions indicate that the investment costs for diffused air systems with a comparable aeration efficiency (evenly distributed diffusers) are higher than for surface aeration systems in general. The power costs should be of the same order to magnitude but maintenance costs for the diffuser cleaning and replacement of diffused air systems may be higher. The critical points of surface aerators are the gears and the bearings. However, experience indicates that surface aerators with well-designed (safety factor 2.0 to 2.5) gears can be operated without malfunction for 15 years or more.

## Vertical shaft or horizontal axis surface aerators?

Mechanically, the difference between the two surface aeration systems is that the vertical shaft aerator receives axial thrusts and radial forces while the horizontal axis aerator is forced radial only. All the resulting forces are transfered to the gear unit of vertical shaft aerators. At horizontal axis aerators the forces are distributed both on the gear and the bearing. For the same power transmission it may be assumed that the forces on the gear of horizontal axis aerators are smaller than those of vertical shaft aerators.

From the point of view of tank construction costs the differences are small. The head ends of the circulation tanks for horizontal aerators are a little more elaborate but on the other hand the cone aerators need bridges or supports for the aerators. The areas to be covered are of comparable size but the covers for the horizontal rotors may be prefabricated.

High oxygen transfer rates are possible with cone aerators (e.g. Hamburg 220 mg/l/hr). Because of the required spacing of at least 25-30 m (80 to 100 ft) between horizontal rotors and an optimal water depth of 2.5 m the maximum oxygen transfer rate is about 100 mg/l/hr. If higher oxygen transfer rates are required horizontal rotors must not be used.

## Loading rates and process requirements

In order to achieve a high quality effluent with regard to BOD and COD most plants in Germany today are designed for a loading rate of F/M 0.15 kg BOD/kg MLSS/d, at which nitrification will be established. Although today ammonia removal is required only at plants in some larger cities, it has to be considered that regulations may call for ammonia removal in every plant in future.

Once nitrification becomes necessary or is considered it is preferable to remove as much nitrate as possible by denitrification without using external sources of carbon in order to save energy. Two successful denitrification systems are pre-denitrification (9) and simultaneous denitrification. The latter was first observed in oxidation ditches (10), and the first larger plant with simultaneous denitrification was Vienna Blumental (11).

Simultaneous denitrification is possible in any completely mixed or circulation tank. The secret of the process is a proper control of the oxygen transfer. It has to be high enough for nitrification and low enough to establish anoxic conditions at certain times or in certain parts of the aeration tank. Simultaneous denitrification is superior to pre-denitrification because no tank separation is needed in order to distinguish between anoxic and oxic tankage. In addition it is more flexible because during periods of higher loads the denitrification rate increases and consequently the anoxic volume or period can be smaller thus leaving more volume for nitrification; the opposite holds for periods with low loads.

Anoxic zones are established in circulation tanks with mammoth rotors or Carrousel tanks in between the aerators. The length of the anoxic zones can be controlled by switching aerators on and off.

By proper control of aeration it might be possible to obtain anoxic zones in completely mixed tanks (for example with cone aerators in the bottom area); another solution is to periodically reduce aeration in order to achieve anoxic periods. It may be necessary to install some mixing device, e.g. a propeller, in order to sustain mixing during periods of reduced aeration at least in smaller plants.

In order to make use of the good flexibility of simultaneous denitrification aeration control is necessary. At the Vienna-Blumental plant the rotors were timer operated at first. Later a continuous respirometer was developed (12) which was specially adapted to the process. It is still in operation and works successfully.

It is thought that the nitrate concentration in the aeration tank would be a better control parameter. First nitrate probes were tested; the results, however, were not promissing. Next a continuous photometric $NO_3$-$NO_2$-analyzer was assembled. The basic problem was to obtain a continuous sample from the aeration tank without suspended solids. After trying different filters without success, the problem was solved by use of an „Ultra-Filtration-Tube" from Kent-BBC. The flow diagram of the set-up is shown in Fig. 9. The results on Fig. 10 indicate that it is possible to achieve a very low nitrate concentration in the effluent.

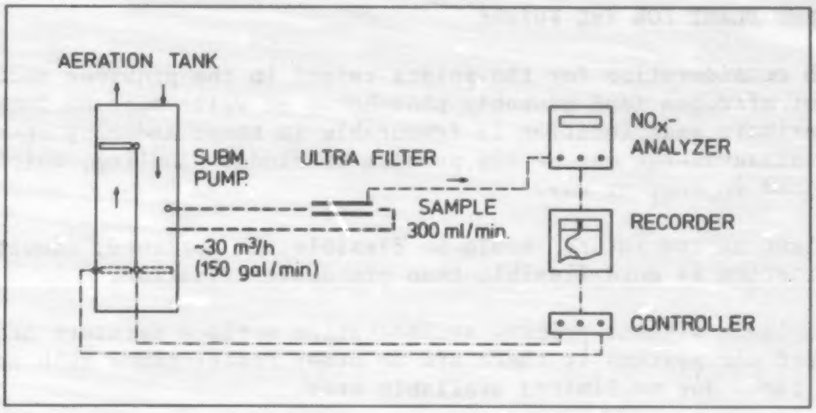
AERATION TANK
SUBM. PUMP
ULTRA FILTER
SAMPLE 300 ml/min.
~30 $m^3/h$ (150 gal/min)
$NO_x^-$ ANALYZER
RECORDER
CONTROLLER

**Figure 9. Aeration control for simultaneous denitrification**

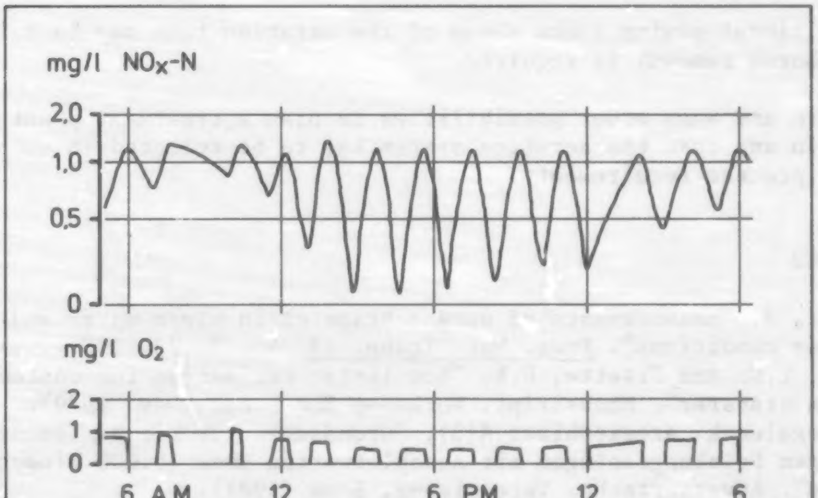

**Figure 10. Nitrate concentrations and D.O. in an aeration tank with controled aeration**

Phosphorus is another source of concern today. It is known that the phosphorus concentration of domestic wastewater can be reduced considerably solely by microbial activity. A mixing tank for wastewater and return sludge is needed ahead of the full denitrifying plant.

In many cases operating without primary sedimentation solves the problem of bulking sludge due to filamentous bacterial growth. Mixing of the wastewater and the return sludge ahead of the operation tank as for phosphorus removal also improves a settle ability of the mixed liquor.

## A TREATMENT PLANT FOR THE FUTURE

With consideration for the points raised in the previous section, the problem of nitrogen (and probably phosphorus as well) must be faced. A plant without primary sedimentation is favourable in these and many other respects. The only disadvantage may be the problem of sludge handling, which may be circumvented in another way.

A plant of the future should be flexible. As outlined, simultaneous denitrification is more flexible than pre-denitrification.

For plants without primary sedimentation surface aerators are superior to diffused air systems if there are no other restrictions such as the need for deep tanks due to limited available area.

Circulation tanks are ideal for simultaneous denitrification because of the sawlike oxygen profile. As parameter for the control of the aeration the nitrate concentration of the mixed liquor is usefull.

Additional mixing tanks ahead of the aeration tank may be built later if phosphorus removal is required.

There are many other possibilities to plan a treatment plant for the future. In any case the aeration system has to be selected in accordance with the process requirements.

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# IMPROVEMENTS OF AIR DIFFUSION SYSTEMS APPLIED IN THE NETHERLANDS

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## ABSTRACT

Field tests with water on improved (uniform arrangement of diffusors or horizontal flow) air diffusion systems are reported and evaluated. Both improvements reduce the upward velocity of the water above the diffusors. Depending on this reduction, 4.5 to 7.7% oxygen absorption per m diffuser depth are obtained and 2.3 to 4.3 kg $O_2$/kWh oxygenation efficiency. Extreme efficiencies are mainly caused by low (40%) or high (60%) motor/blower efficiencies.

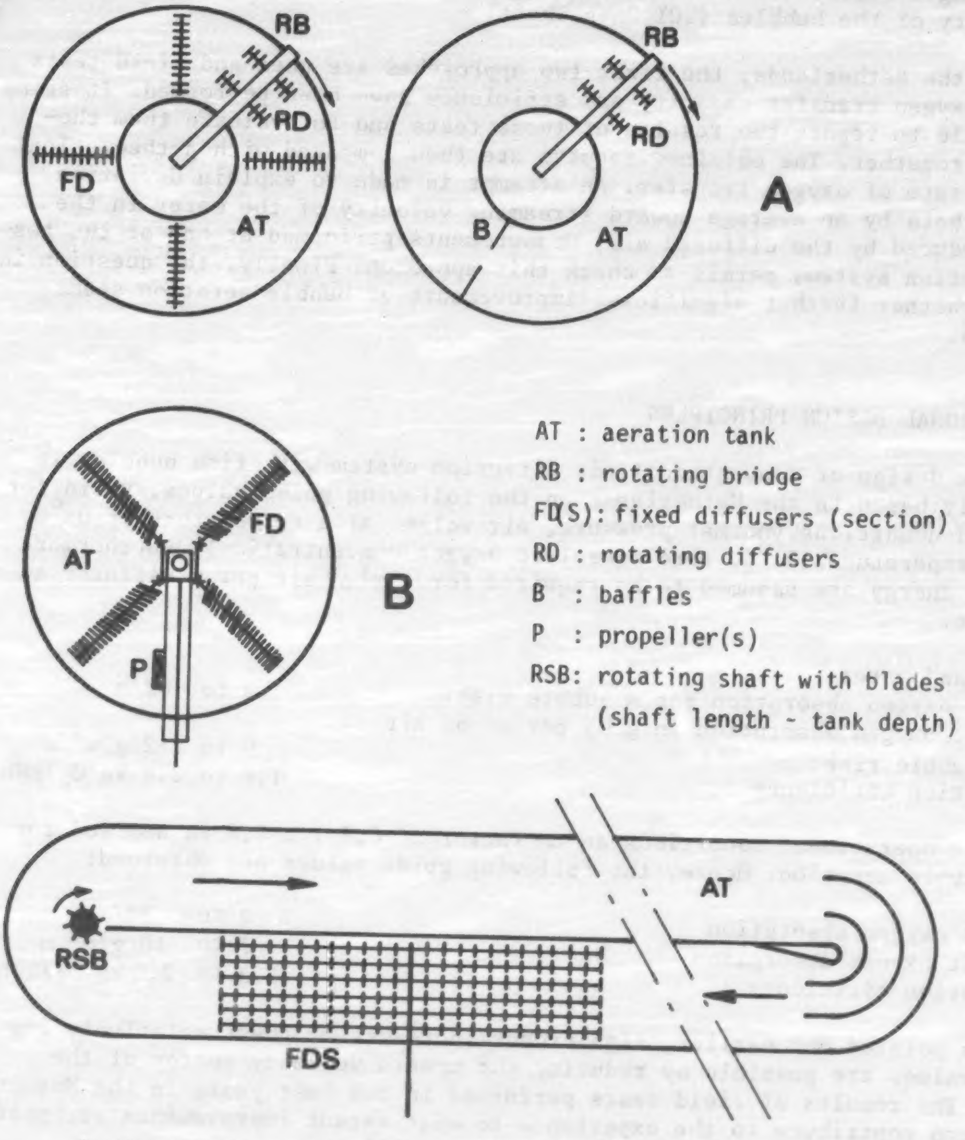
Upward velocities are measured in one system. Their influence on oxygen transfer is modelled and applied to the other systems. Model results allow to assess whether further improvements of oxygen transfer and efficiency by changing the system design are possible.

## INTRODUCTION

In the Netherlands, roughly 70% of the municipal wastewater is treated by the activated sludge process. 30% of the total treatment capacity is represented by extremely low loaded plant (oxidation ditch principle) and the majority of the remaining 40% is designed for nitrification by applying low sludge loading rates also. The implications with respect to aeration are:
- relatively low densities of oxygen transfer (g $O_2/m^3$ tank volume.h) are required which may cause mixing problems;
- relatively large amounts of oxygen have to be transferred per unit mass of BOD removal. This would mean that improvements of the areation system could yield considerable savings at increasing energy prices.

Improvements of the air diffusion technology, but also more stringent (legal) requirement with respect to noise and odor problems, have stimulated the application of fine bubble aeration again and at the same time reduced further expansion of mechanical aeration.

The technological improvements are essentially based on the "new development in diffused air aeration" as put forward by Pasveer and Sweeris in 1965 (1): by special means swift upward flow velocities in the aeration tank above the diffusors have to be prevented because they significantly reduce the retention time of the bubbles in the water and hence the total interfacial area. Practical means to achieve this goal are:
- a uniform arrangement of the diffusors all over the tank bottom (no spiral flow);
- creating horizontal streaming velocities in a circular tank by a rotating bridge (counter-currant aeration), by propellors mounted within the tank (e.g. Roto-Flow aeration), or by a vane-column mounted in a circuit-like


RB
RD
FD
AT
RB
RD
B
AT
A
AT
FD
P
B
AT : aeration tank
RB : rotating bridge
FD(S): fixed diffusers (section)
RD : rotating diffusers
B : baffles
P : propeller(s)
RSB: rotating shaft with blades
(shaft length - tank depth)
AT
RSB
FDS
C

A: Counter-Current Aeration with and without Fixed Diffusers

B: Roto-Flow Aeration

C: Landox Aeration

Fig. 1: Horizontal Flow Air Diffusion System

tank (see Figure 1).
- creating a downward flow water velocity, somewhat larger than the rising velocity of the bubbles (ICI Deep Shaft).

In the Netherlands, the first two approaches are used and field tests on the oxygen transfer capacity and efficiency have been performed. It seems worthwhile to report the results of these tests and to evaluate them thoroughly together. The obtained results are then compared with a theoretical maximum rate of oxygen transfer. An attempt is made to explain deviation between both by an average upward streaming velocity of the water in the tank, induced by the diffused air. Measurements performed at one of the tested aeration systems permit to check this approach. Finally, the question is raised whether further significant improvements of bubble aeration seems possible.

## CONVENTIONAL DESIGN PRINCIPLES

The design of conventional air diffusion system with fine bubbles is generally based in the Netherlands on the following guide values, valid for standard conditions (normal pressure, air volume at a temperature of 0 °C, water temperature 10 °C, zero dissolved oxygen concentration). Furthermore 5 Wh of energy are assumed to be required for 1 m$^3$ of air per m diffusor submergence.

In "clean" water:

| | |
|---|---|
| percent oxygen absorption per m bubble rise: | 3 to 4%/m |
| specific oxygen absorption in g $O_2$ per m$^3$ of air per m bubble rise: | 9 to 12 g/m$^3$.m |
| oxygenation efficiency | 1.8 to 2.4 kg $O_2$/kWh |

For operational conditions an α- factor of 0,7 ... 0,8 is applied for fine bubble aeration. Hence, the following guide values are obtained:

| | |
|---|---|
| percent oxygen absorption | 2 to 3%/m |
| specific oxygen absorption | 7 to 10 g/m$^3$.m |
| oxygenation efficiency | 1.3 to 2.0 kg $O_2$/kWh |

As pointed out earlier, significant increases of these relatively low guide values are possible by reducing the upward velocity vector of the water. The results of field tests performed in the last years in the Netherlands can contribute to the experience to what extent improvements are possible at present.

## OXYGENATION CAPACITY DETERMINATIONS IN WATER UNDER FIELD CONDITIONS

At a number of activated sludge treatment plants oxygenation capacity determination have been performed prior to starting operations. Much of the results has been published elsewhere, but they never have been evaluated together.

| Table 1 Tested Aeration Systems | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| Nr. | Plant | Tank dimensions (m) | | | | diffusers and arrangement | | | | Ref. |
| | | w * | d * | l = | V ($m^3$) | $d_s$ | type | density | arr. | |
| 1 | Steenwijk | 6.75 | 4.00 | 6.30 | 170 | 3.80 | Act.Sl.Dome | $4.7/m^2$ | u | 2 |
| 2 | " | | same | | | | same | $2.8/m^2$ | u-cc | 2 |
| 3 | Amsterdam | 4.40 | 3.70 | 33.00 | 535 | 3.35 | Degremont | $4.0/m^2$ | u | 3 |
| 4 | " | | same | | | | domes | $1.6/m^2$ | u-c | 3 |
| 5 | " | 4.40 | 4.13 | 33.00 | 600 | 3.85 | same | $1.7/m^2$ | u-cc | 4 |
| 6 | " | | same | | | 4.00 | Brandol 60 | $1.4/m^2$ | u | 4 |
| 7 | " | | same | | | 4.00 | NokiaHKP500 | $1.4/m^2$ | u | 4 |
| 8 | " | | same | | | 4.00 | Schuler Fle<br>Xolith V400 | $1.4/m^2$ | u | 4 |
| 9 | Nieuw-Lekkerland | Counter-Current<br>diam. 25 and 36 m<br>depth 3.20 m | | | 1750 | 3.00 | Brandol 60<br>45% rotat.<br>55% fixed | $.13/m^3$ | $h_f$ | 5 |
| 10 | Hulst | Roto-Flow<br>diam. 42 m<br>depth 4.50 m | | | 5961 | 4.00 | Cellpox | $.36/m^2$ | $h_f$ | 6 |

$d_s$ = depth of diffusor submergence

u = uniform arrangement

u - c(c) = like u, but with some (more intensive) circular motion

$h_f$ = horizontal flow

| Table 2 Test Specifications | | | |
|---|---|---|---|
| Nr. | T-Range | specific air flows | |
| | °C | $m_a^3/m_w^3.h$ | $m_a^3$/h.diff. |
| 1 | 6.0 - 9.0 | 0.95 - 5.73 | 0.81 - 4.88 |
| 2 | 6.0 - 9.0 | 0.51 - 3.76 | 0.72 - 5.33 |
| 3 | 16.4 - 19.6 | 1.07 - 5.24 | 1.01 - 4.85 |
| 4 | 15.8 - 19.5 | 1.10 - 3.00 | 2.56 - 6.95 |
| 5 | 16.5 - 17.5 | 1.45 - 2.42 | 3.30 - 6.00 |
| 6 | 17.0 - 17.5 | 1.43 - 2.35 | 4.00 - 6.70 |
| 7 | 16.5 - 18.0 | 1.44 - 2.17 | 4.20 - 6.30 |
| 8 | 16.0 - 17.0 | 1.39 - 2.29 | 4.00 - 6.70 |
| 9 | 19.5 - 21.0 | 0.21 - 0.42 | 5.13 - 10.2 |
| 10 | 19.0 - 20.0 | 0.11 - 0.37 | 1.34 - 4.40 |

In all cases the oxygenation capacity has been determined by the unsteady state method after deoxygenation with sodium sulfite and cobalt as catalyst. From the course of the oxygen concentration the mass transfer coefficient $K_L.a$ ($h^{-1}$, base e) is determined by applying various techniques and finally the OC by

$$OC = 0.0113.K_L.a.V.f_T.f_c \qquad (kg\ O_2/h) \quad (1)$$

with $V$ = volume of water aerated ($m^3$)

$f_T$ = temperature correction factor, generally based on exponential model with $\theta = 1.018$ to $1.02$

$f_c$ = saturation concentration correction factor = obtained supersaturation factor by test

0.0113 = standard saturation concentration in kg $O_2/m^3$

$K_L$ = "liquid film" coefficient (m/h)

$a$ = specific interfacial area = total bubble area A divided by water volume V.

Other transfer rate parameters to be discussed later, can be derived from the OC or $K_L.a$ and the specific characteristics of the aeration system and the test.
The oxygenation efficiency (kg $O_2$/kWh) is calculated by dividing the OC (kg $O_2$/h) by the gross power (kW) drawn from the distribution net.

## Tested Aeration Systems

In total, tests on 10 aeration systems at 4 plants are reported. For each system between 3 and 14 tests were performed by varying the airflow. Details of the aeration systems are given in table 1: tank dimensions, diffuser type, density and arrangement and also their depth of submergence ($d_s$) are stated. In total, $d_s$ varies between 3.35 m and 4.0 m, the diffuser density between 1.4 and 4.7 per $m^2$ for normal flow conditions and between only 0.13 and 0.36 per $m^2$ for horizontal flow systems (including the rotating diffusers).

## Test Specification

The most important test conditions are specified in table 2 by giving the range of water temperatures (6 to 21 °C) and information on the air flows applied. For better comparison among the systems, only specific rates of air flows are reported. Firstly, the $m^3$ of air diffused per $m^3$ of water per hour is stated, ranging from 1.0 to 5.7 $m^3/m^3.h$ for the normal systems and amounting to only one tenth of these values for the horizontal flow systems. Secondly, the rate per diffuser is given ($m^3$ of air per diffuser per hour), ranging from 0.7 to 10 $m^3$/h.diff for all systems.

## Test Results

The test results are compiled in table 3. The measured oxygenation capacities (eq.1) are converted into specific oxygen transfer measures. Firstly, the range of specific oxygen absorption (g $O_2$ absorbed per $m^3$ of air per m diffuser submergence = bubble rise) is given, followed by the percent oxy-

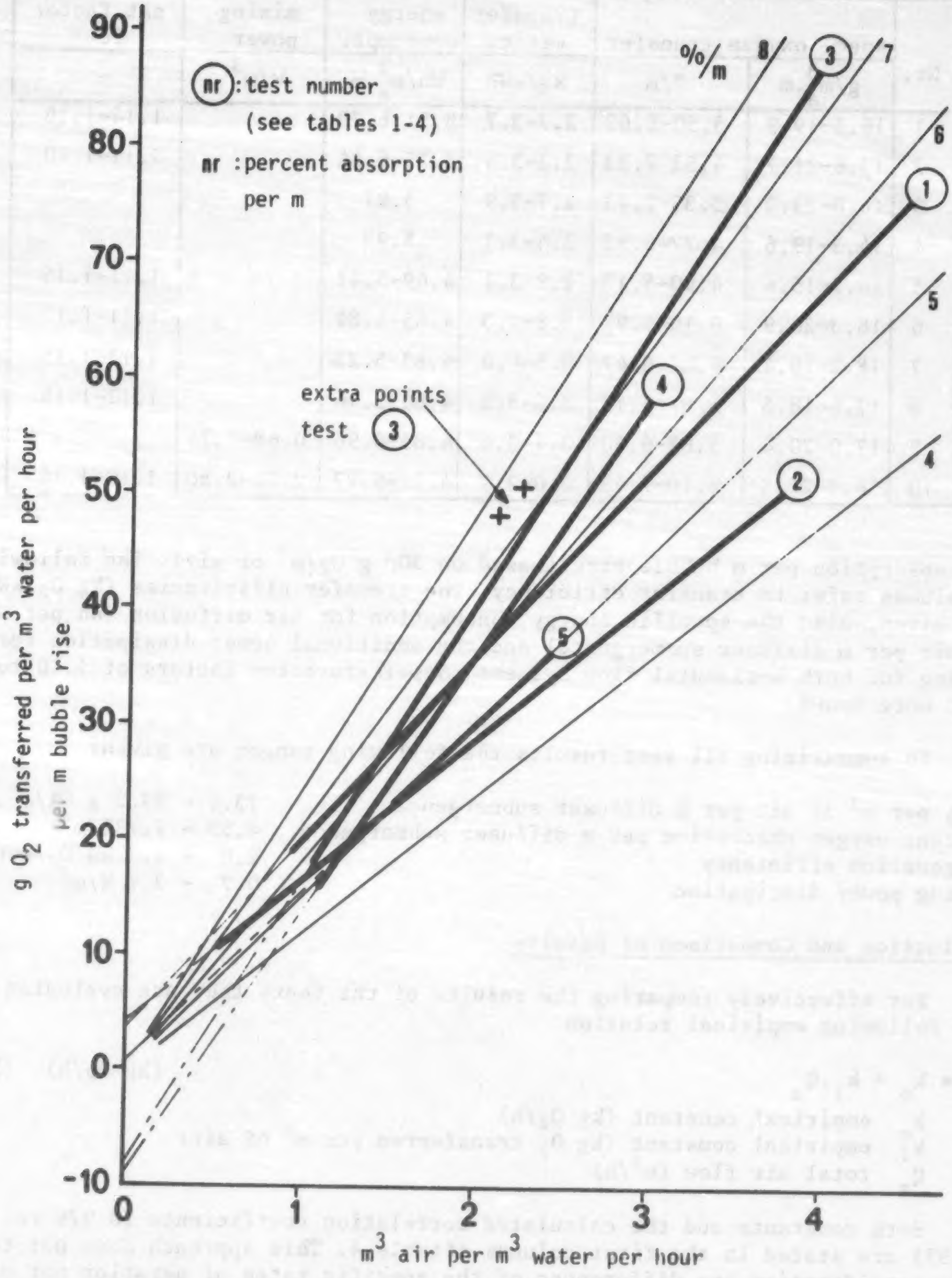


Fig. 2: Standardized Results of Tests 1 to 5

Table 3 Test Results

| Nr. | spec. oxygen transfer |  | transfer effic. | energy consumpt. | mixing power | sat.factor fc |
|---|---|---|---|---|---|---|
|  | $g/m_a^3.m$ | %/m | kg/kWh | $Wh/m_a^3.m$ | $W/m^3$ | - |
| 1 | 16.5-19.9 | 5.50-6.63 | 2.7-3.7 | 5.31-6.73 |  | 1.14-1.18 |
| 2 | 13.6-21.7 | 4.53-7.23 | 2.3-3.4 | 5.25-6.84 |  | 1.11-1.20 |
| 3 | 16.0-23.2 | 5.33-7.73 | 2.7-3.9 | 5.97 |  |  |
| 4 | 14.3-19.6 | 4.77-6.53 | 2.4-3.3 | 5.97 |  |  |
| 5 | 14.7-15.4 | 4.90-5.13 | 2.9-3.1 | 4.69-5.41 |  | 1.11-1.15 |
| 6 | 18.3-20.9 | 6.10-6.97 | 3.8-4.3 | 4.65-4.89 |  | 1.11-1.12 |
| 7 | 18.2-19.4 | 6.07-6.47 | 3.5-4.0 | 4.83-5.23 |  | 1.13-1.15 |
| 8 | 17.6-18.5 | 5.87-6.17 | 3.6-3.8 | 4.60-5.16 |  | 1.12-1.15 |
| 9 | 17.0-20.4 | 5.67-6.80 | 3.1-3.4 | 4.85-4.96 | 0.68-0.74 |  |
| 10 | 18.3-22.3 | 6.10-7.43 | 2.0-2.7 | 4.79-5.27 | 2.78-2.80 | 1.10-1.16 |

gen absorption per m bubble rise (based on 300 g $O_2/m^3$ of air). The following 3 columns refer to transfer efficiency. The transfer efficiencies (kg $O_2$/kWh) are given, also the specific energy consumption for air diffusion (Wh per $m^3$ of air per m diffuser submergence) and the additional power dissipation for mixing for both horizontal flow systems. Supersaturation factors of 1.10 to 1.20 were found.

In summarizing all test results the following ranges are given:

| | |
|---|---|
| g $O_2$ per $m^3$ of air per m diffuser submergence | 13.6 - 23.2 g $O_2/m^3$.m |
| percent oxygen absorption per m diffuser submergence | 4.53 - 7.73%/m |
| oxygenation efficiency | 2.0 - 4.3 kg $O_2$/kWh |
| mixing power dissipation | 0.7 - 2.8 $W/m^3$ |

## Evaluation and Comparison of Results

For effectively comparing the results of the tests they are evaluated by the following empirical relation

$$OC = k_o + k_1 . Q_a \qquad (kg\ O_2/h) \quad (2)$$

$k_o$ empirical constant (kg $O_2$/h)
$k_1$ empirical constant (kg $O_2$ transferred per $m^3$ of air)
$Q_a$ total air flow ($m^3$/h)

Both constants and the calculated correlation coefficients (0.979 to 0.999) are stated in the first columns oftable 4. This approach does not take into consideration any differences of the specific rates of aeration nor of different depths of diffuser submergence. Therefore, equation 2 is divided by the water volume V ($m^3$) and the diffuser depth $d_s$, to obtain the specific

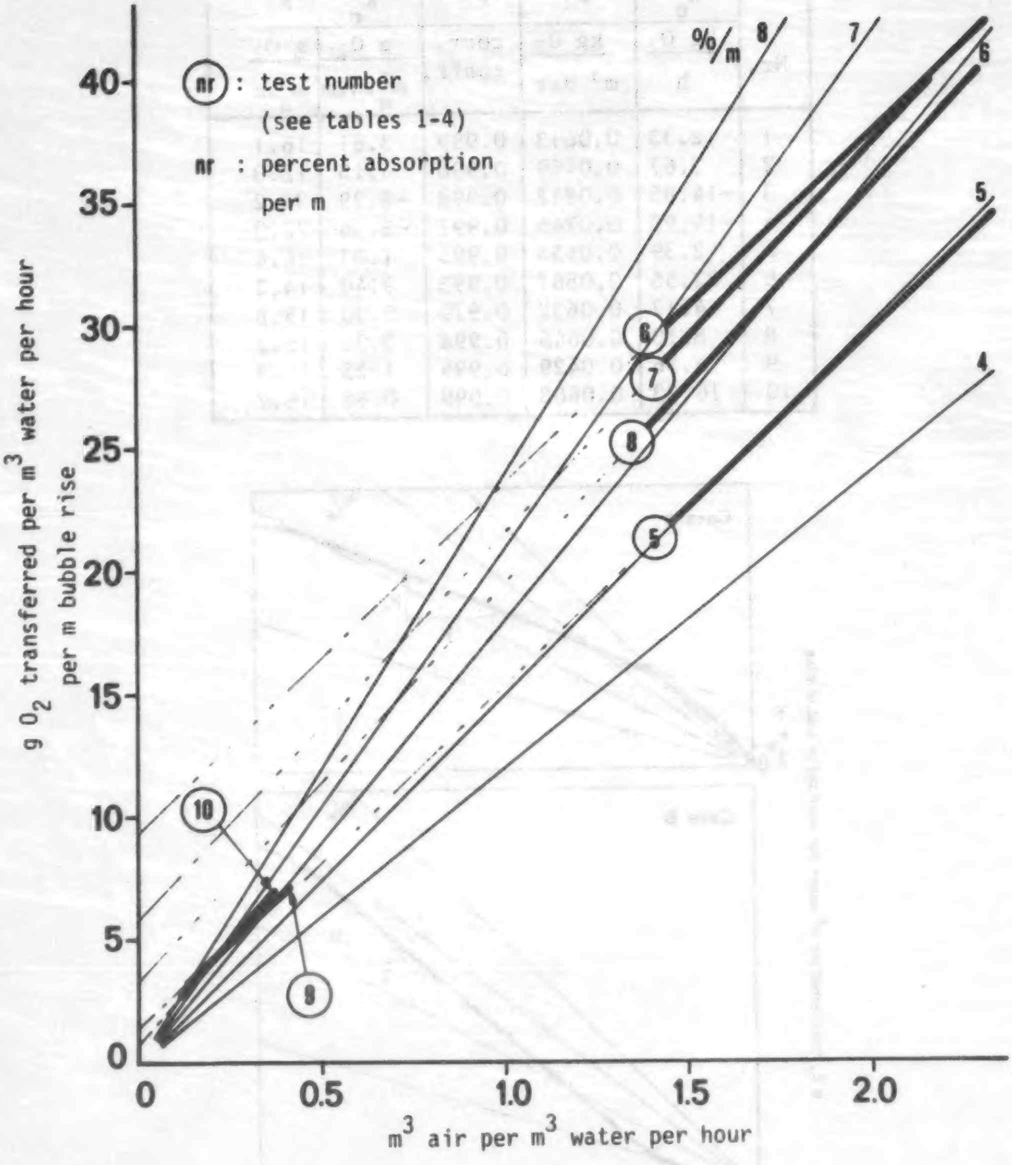


Fig. 3: Standardized Results of Tests 5 to 10

Table 4 Standardized Evaluation of Tests

| Nr. | $k_o$ | $k_1$ | r | $K_o$ | $K_1$ |
|---|---|---|---|---|---|
| | $\frac{kg\ O_2}{h}$ | $\frac{kg\ O_2}{m^3\ air}$ | corr. coeff. | $\frac{g\ O_2}{m_w^3.h.m}$ | $\frac{g\ O_2}{m_a^3.m}$ |
| 1 | 2.33 | 0.0613 | 0.999 | 3.61 | 16.1 |
| 2 | 2.67 | 0.0469 | 0.998 | 4.13 | 12.3 |
| 3 | -14.85 | 0.0812 | 0.998 | -8.29 | 24.2 |
| 4 | -14.98 | 0.0745 | 0.997 | -8.36 | 22.3 |
| 5 | 2.39 | 0.0555 | 0.995 | 1.03 | 14.4 |
| 6 | 22.56 | 0.0567 | 0.993 | 9.40 | 14.2 |
| 7 | 14.17 | 0.0632 | 0.979 | 5.90 | 15.8 |
| 8 | 8.10 | 0.0646 | 0.998 | 3.38 | 16.2 |
| 9 | 6.56 | 0.0429 | 0.999 | 1.25 | 14.3 |
| 10 | 16.33 | 0.0668 | 0.999 | 0.68 | 16.7 |

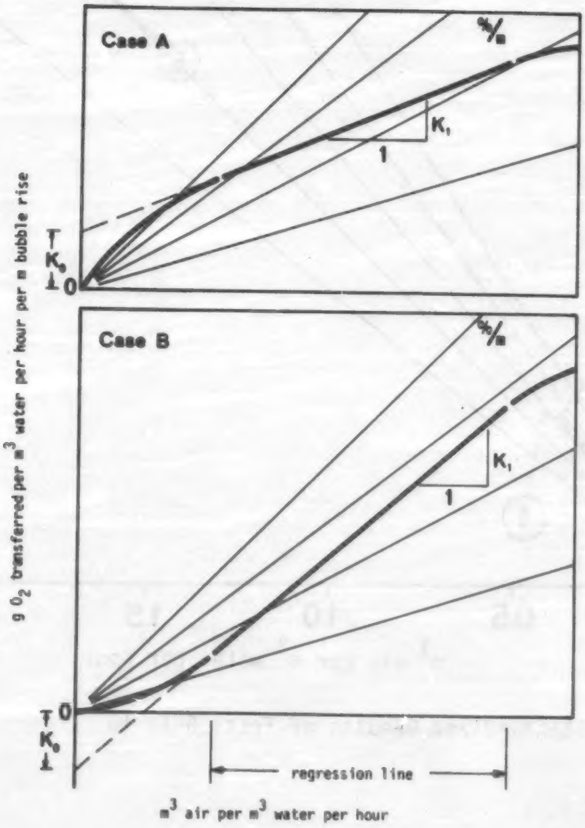

Case A
%/m
K₁
1
K₀
0
Case B
%/m
K₁
1
0
K₀
g O₂ transferred per m³ water per hour per m bubble rise
regression line
m³ air per m³ water per hour

Fig. 4: Basic Interpretation of Transfer Regression Line

oxygenation capacity SOC ($g\ O_2$ transferred per $m^3$ of water per hour for one m of bubble rise):

$$SOC = \frac{1000.k_o}{V.d_s} + \frac{1000.k_1}{d_s}.\frac{Q_a}{V} \quad \frac{g\ O_2}{m^3\ water.h.m} \qquad (3)$$

which is rewritten as

$$SOC = K_o + K_1.q_a \quad \frac{g\ O_2}{m^3\ water.h.m} \qquad (4)$$

$$K_o = \frac{1000.k_o}{V.d_s} \left(\frac{g\ O_2}{m^3\ water.h.m}\right) = \text{empirical constant}$$

$$K_1 = \frac{1000.k_1}{d_s} \left(\frac{g\ O_2}{m^3\ air.m}\right) = \text{empirical constant}$$

$q_a$ = specific rate of aeration ($m^3$ air/$m^3$ water.h)

The standardized constants $K_o$ and $K_1$ are also given in table 4. The correlation coefficients hold for this relation also. The results are graphically represented in figures 2 (test 1 to 5) and 3 (test 5 to 10). The graphs contain also lines of constant percent oxygen absorption per m diffuser depth, which allows to easily judge the performance of the aeration system.

## Discussion of Regression Equation

The intersection $K_o$ of equation 4, being mathematically independent of the specific air flow $q_a$, varies over a considerably large range (-8,4 to + 9.4), It is not possible to interprete $K_o$ as the contribution of the tank water surface to oxygen transfer, because (a) this would include negative contributions (tests 3 and 4), (b) some contributions (55 to 155 $g\ O_2/m^2.h$) are unrealistically high (tests 1, 2, 6, 7, 8). Only test 5 and both horizontal flow systems (9 and 10) would give reasonable values ranging from 11.8 to 16.5 $g\ O_2/m^2.h$.

A more reasonable explanation of high positive values of $K_o$ can be given via a relation between the SOC and the specific air flow $q_a$ over a wider range than applied in the reported tests. This assumed relation is represented in figure 4A. At very low air flows the produced bubbles are very small, upward velocities of the water almost negligible. As a consequence of both effects, the interfacial bubble area is relatively large and so is the percent oxygen absorption. At increasing specific air flow $q_a$ both effects become less favorable and the percent oxygen absorption is decreasing. After this transition the range of a straight line relationship between SOC and $q_a$ is reached. Exceeding this range a smaller increase of the SOC is conceived of than predicted by the straight line.

Within this reasoning a third effect has not been taken into account, namely the decrease of oxygen in the air bubbles during transfer, its effect on the saturation concentration and hence on the rate of transfer. This effect is obviously greatest at high percent oxygen absorptions, i.e. at low specific air flows $q_a$. If this effect is of significance, then an increase of $q_a$ will increase the mass of oxygen transferred per m$^3$ of air (per m). But this latter increase will be such, that the percent oxygen absorption is decreasing. Thereby the oxygen content in the air bubbles is higher than at lower rates of airflow and so are the saturation concentration and the rate of transfer. If, with increasing $q_a$, the percent oxygen absorption remains (almost) constant ($K_o$ relatively small) or increases even ($K_o$ negative), then effects other than oxygen depletion in the diffused air are responsible. Since the bubble size increases with increasing $q_a$, the only explanation for a negative $K_o$ can be an assumption about the mixing conditions becoming more favorable as the air flow increases (see figure 4B). Such changes cannot easily be conceived of and therefore also this does not explain a negative $K_o$-value satisfactorily.

Interestingly enough, 4 runs with the aeration system 3 ($K_o = - 8.3$!) gave at some 2.2 m$^3$ air per m$^3$ water per hour exceptionally high SOC-values of some 50 g $O_2$/m$^3$ water per hour per m bubble rise (7.6%/m). These 4 measurements (see "extra points test 3") do not belong to the regression population of line 3 in the statistical interpretation.

The slope of the regression lines $K_1$ varies from 12.3 to 24.2 g $O_2$/m$^3$ air per m. Converted to percent oxygen absorption this would mean 4.1 to 8.1% per m. But these values have to be interpreted together with the intersection $K_o$. The actual range of specific oxygen transfer can be seen in table 3 and in figures 2 and 3 and goes from 13.6 to 23.2 g $O_2$/m$^3$.m and from 4.5 to 7.7%.

There is a tendency that large values of the slope go along with low values for the intersection (negative correlation). For all 10 test a significant (P > 99%) correlation coefficient (r = - 0.862) is calculated, mainly due to large positive and negative values of $K_o$. Omitting the negative $K_o$, the correlation (r = - 0,209) remains negative, but becomes insignificant.

## Oxygenation Efficiency

The oxygenation efficiency OE (kg $O_2$/kWh) of the aeration systems can be calculated by dividing the following 2 parameters: (a) the specific oxygen transfer rate (g $O_2$ per m$^3$ air per m diffuser depth) over (b) the specific energy consumption in Wh (= 3.6 kJ) per m$^3$ of diffused air per m diffuser depth. The first factor varies over a considerable range (see table 3), whereas the second does not change very much. Measured values for root blowers (test 5 to 10) show a range of only 4.5 to 5.4 Wh/m$^3$.m. The value of 6.0 of tests 3 and 4 has been assumed from earlier measurements. The aeration systems 1 and 2 had a centrifugal blower showing low efficiency and correspondingly high specific energy consumption (6.8 Wh/m$^3$.m) when throttled to obtain low air flows. Without throttling a value of 5.3 Wh/m$^3$.m is obtained.

For horizontal flow systems the energy used by the seperate water propulsion device has to be added to the above estimate. The specific power dissipation (W/m$^3$) is a reliable incidation of this element of energy comsumption. Since it is (almost) independent of the air flow, its share is relatively high at low air flows and oxygen transfer rates (and vice versa). System 9 dissipates only 0.7 W/m$^3$ and shows OE-values above 3 kg $O_2$/kWh, therefore;system 10, however, dissipating 2.8 W/m$^3$ with the propellors, does not exceed 2.7 kg $O_2$/kWh.

The above considerations allow to explain extremely low OE values of the tests as follows:

- the lower values of tests 2 and 4 are caused by low transfer rates and relative high specific energy consumption;
- the lower values of test 1 and 3 can be explained by high specific energy consumption;
- the low values of test 10 are due to the high power dissipation for mixing.

Avoiding above drawbacks it seems possible to reach at least 3 kg $O_2$/kWh in water and 2.2 kg $O_2$/kWh under operational conditions for systems with small upward velocities. Maximum values of 4 kg $O_2$/kWh in water and of 3 kg $O_2$/kWh under operational conditions appear realistic.

## INTERPRETATION OF RESULTS

The uniform arrangement of diffusers in relatively deep tanks (4 m) to prevent (double) spiral flow and the development of horizontal flow systems aim at reducing vertical streaming velocities above the diffusers and provide at the same time enough mixing intensity to keep the activated sludge solids in suspension. In this section an attempt is made to estimate on theoretical grounds to what extent such a reduction is accomplished in the tested aeration systems. Measurements of the vertical velocity component above the diffusers in one of the systems, knowledge of the bubble size of that system and a simple estimate of the effect of oxygen decrease in the diffused air allow to check the theoretical approach.

### Theoretical Background

The approach is based on the "penetration theory" proposed by Higbie (7) and worked out for bubble aeration by Pasveer (8). According to this theory the "liquid film coefficient" or better liquid transfer coefficient $K_L$ is given by

$$K_L = 2\sqrt{\frac{D.v_s}{\pi.d_B}} \qquad \text{m/s} \qquad (5)$$

$D$ = diffusion coefficient of oxygen in water
at 10 °C $D = 1.39.10^{-9}$ m²/s $= 5.00.10^{-6}$ m²/h

$v_s$ = "slip" velocity of bubbles in water

$d_B$ = diameter of "spherical" bubbles

The slip velocity of bubbles of 2 to 7 mm diameter in water is constant (9) amounting to $v_s = 0.23$ m/s and is independent of temperature (10). In filtered or pure water there is a decrease with increasing bubble size within the above range (10):

$$v_s = \sqrt{\frac{3.64.\sigma}{\rho.d_B}} \qquad \text{m/s} \qquad (6)$$

$\sigma$ = surface tension of water at 10 °C $\sigma = 0.0742$ N/m

$\rho$ = density of water
$v_s$ (at 10 °C) = $0.0164\sqrt{d_B}$ m/s

The liquid transfer coefficient can therefore be estimated for water when the bubble size is known. For the Nokia-diffuser (test 7) the bubble size was (11)

$$d_B = 0.080 . q_d + 2.9 \text{ mm} \tag{7}$$
$q_d$ = air flow m$^3$/h.diffuser

This information allows to calculate the average liquid transfer coefficient as a function of the air flow $q_d$.

In order to apply equation 1 for theoretical estimation of the OC, the specific interfacial area of the bubbles $a = A/V$ has to be known. The total area depends on (1) the total air flow $Q_a$ (m$^3$/h), (2) the bubble size (diameter of spherical bubbles), (3) the retention time of the produced bubbles in the water $\tau_B$ which equals the diffuser depth $d_s$ divided by the bubble rise velocity in the tank $v_r = v_s + v_v$.
For spherical bubbles we get

$$A = \frac{6 . Q_a . \tau_B}{d_B} = \frac{6 . Q_a . d_s}{d_B . (v_s + v_v)} = \frac{6 . Q_a . d_s}{d_B . f_v . v_s} \tag{8}$$

$f_v$ = bubble velocity factor
$= \dfrac{\text{rising velocity of bubbles above diffuser}}{\text{slip velocity of bubbles in quiescent water}}$
= 2 to 3 in spiral flow tanks (8)

Combining equations 1, 5, and 8 yields with $a = A/V$

$$OC = 0.1356 \cdot \sqrt{\frac{D}{\pi . v_s . d_B^3}} . Q_a . d_s . \frac{f_T . f_c}{f_v} \text{ kg } O_2/\text{h} \tag{9}$$

The oxygen supplied to the system by the total air flow $Q_a$ amounts to $0.3 . Q_a$ kg $O_2$/h. The fraction of oxygen absorbed per m of diffusor depth p' is then obtained by dividing equation 9 by $0.3 . Q_a . d_s$ to give

$$p_1' = 0.452 \cdot \sqrt{\frac{D}{\pi . v_s . d_B^3}} . \frac{f_T . f_c}{f_v} \qquad \text{m}^{-1} \tag{10}$$

Inserting a constant $v_s = 0.23$ m/s and the stated numerical values into equation 10 gives for $d_B$ in mm

$$p' = \frac{0.627 . f_T . f_c}{f_v . d_B^{1.5} (\text{mm})} \qquad \text{m}^{-1} \tag{11}$$

Assuming filtered or pure water conditions, equation 6 is inserted into 10 to give with the indicated numerical values for $d_B$ in mm

$$p_2' = \frac{0.417 . f_T . f_c}{f_v . d_B^{1.25} (\text{mm})} \qquad \text{m}^{-1} \tag{12}$$

Multiplication of equations 11 and 12 with 100 would give the percent oxygen absorption per m diffuser depth $p_1$ and $p_2$. Maximum values of p at 10 °C ($f_T = 1$) are obtained for conditions of quiescent water ($f_v = 1$). $p_1$ is plotted for a typical oversaturation factor of 1.15 for bubble sizes of 2 to 7 mm in figure 5. For this range, percent oxygen absorptions decreasing from 25%/m to 4%/m are calculated.

The greater percentage absorption, the more the transfer limiting influence of oxygen depletion in the diffused air becomes significant. The above theoretical approach by equations 11 and 12 should be corrected for this influence, therefore. It can be shown - along similar reasoning as presented in (12) - that the correction factor for oxygen depletion $f_o$ can be estimated by

$$f_o = \frac{1}{1 + \frac{0.531 . d_s}{f_c} . p'} \qquad (13)$$

The corrected absorption $p_c'$ can then be estimated by

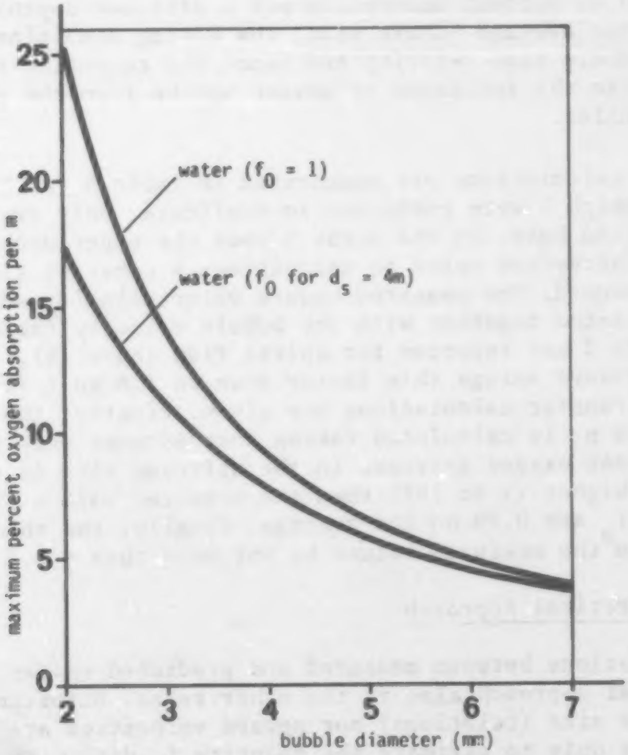


Fig. 5: Maximum Percent Oxygen Absorption from Spherical Bubbles at $f_c = 1.15$ in Quiescent Water

$$p'_{1,c} = f_o \cdot p'_1 \qquad p'_{2,c} = f_o \cdot p'_2 \qquad m^{-1} \qquad (14)$$

$p_{1,c}$-values (%/m) are plotted for $f_c = 1.15$ and $d_s = 4$ m in figure 5. A comparison with the not corrected $p_1$-values shows the significant influence of oxygen depletion in the diffused air, especially at high absorption, i.e. with small bubbles. The corrected percent absorptions per m decrease in the cited bubble range from 17 (was 25) to 3,3 (was 4) %/m.

## Verification of Theoretical Approach

For test 7 not only the bubble size was known as a function of the diffuser loading ($m^3$ air/h.diff.) as given by equation 7, but also the upward velocity $v_v$ of the water half-way above the diffusers was measured as a function of the specific air flow $q_a$ ($m^3$ air/$m^3$ water per hour). The relation was (11) generalized (r = 0.929) by

$$v_v = 0.121 . \sqrt{q_a} - 0.01 \qquad m/s \qquad (15)$$

Hence all information required to theoretically calculate the oxygen transfer (expressed as percent absorption per m diffuser depth) is available. In this approach, the average bubble size, the mixing conditions in the tank as affecting the bubble size velocity and hence its retention time are taken into account and also the influence of oxygen uptake from the diffused air on the rate of transfer.

The pertinent calculations are summarized in table 5 for 6 aeration test runs (system 7) of which 3 were performed in duplicate. Only averages of every duplicate are reported here. In the first 5 rows the experimental data are given. The 4 rows thereafter refer to velocities. A constant slip velocity of $v_s = 0.23$ m/s is assumed. The measured upward water velocity and the bubble rise velocity are stated together with the bubble velocity factor $f_v$. Whereas $f_v$-values of 2 to 3 are reported for spiral flow tanks (8), the uniform arrangement of diffusers brings this factor down to 1.6 to 1.7. In the last 4 rows the oxygen transfer calculations are given. Firstly, the percent oxygen absorption per m $p_1$ is calculated taking into account only the measured upward velocities, not oxygen decrease in the diffused air. As expected, this prediction in much higher (9 to 19%) than the measured values. The oxygen correction factors $f_o$ are 0.88 on the average. Finally, the theoretical predictions differ from the measured values by not more than - 3.5 to + 4.1%.

## Application of Theoretical Approach

The small deviations between measured and predicted values encourage to apply the theoretical approach also to the other tests. Unfortunately, however, neither bubble size (relations) nor upward velocities are known. Hence, this approach allows only to estimate the quantity $f_v . d_B^{1,5}$ with reasonable accuracy along the following lines.

The measured oxygen absorption per m is assumed to be equal to the corrected theoretical absorption value $p'_{1,c}$. The uncorrected theoretical absorption $p'_1$ is then calculated (inversion of eq. 13 and 14) by

| Table 5 Theoretical Transfer Calculations | | | | |
|---|---|---|---|---|
| Parameter | Dimension | Average value of run | | |
| | | 1+2 | 2+3 | 4+5 |
| Data | | | | |
| diff.depth $d_s$ | m | 4.05 | 4.05 | 4.05 |
| spec. air flow $q_a$ | $m_a^3/m_w^3.h$ | 2.157 | 1.803 | 1.438 |
| spec. air flow $q_d$ | $m_a^3$/h.diff | 6.25 | 5.25 | 4.20 |
| sat. factor $f_c$ | - | 1.138 | 1.135 | 1.141 |
| perc. oxyg. abs. p | %/m | 6.065 | 6.37 | 6.48 |
| bubble diam. (eq.7) | mm | 3.40 | 3.32 | 3.24 |
| Velocities | | | | |
| slip velocity $v_s$ | m/s | 0.23 | 0.23 | 0.23 |
| upw. veloc.$v_v$ (eq.15) | m/s | 0.168 | 0.152 | 0.135 |
| bubble vel.$v_r=v_s+v_v$ | m/s | 0.398 | 0.382 | 0.365 |
| vel.factor $f_v=v_r/v_s$ | - | 1.729 | 1.663 | 1.587 |
| Transfer calculat. | | | | |
| $p_l$(eq.11)uncorr. | %/m | 6.583 | 7.074 | 7.730 |
| corr. fact.$f_o$(eq.13) | - | 0.889 | 0.882 | 0.873 |
| $p_{l,c}$(eq.14)corrected | %/m | 5.852 | 6.235 | 6.747 |
| perc.deviat.from measured p | % | -3.51 | -2.13 | +4.12 |

$$p_l' = \frac{p_{l,c}'}{1 - \frac{0.531.d_s.p_{l,c}'}{f_c}} \quad m^{-1} \tag{16}$$

The above product can be estimated with this value by means of the rearranged equation 11:

$$f_v.d_B^{1,5}\,(mm) = \frac{0.627.f_T.f_c}{p_l'} \tag{17}$$

The required value of $p_{l,c}'$ (eq. 16) is calculated by means of equation 2 based on the average air flow of each test. The results of this approach are summarized in table 6 by stating

- the product $f_v.d_B^{1,5}$(mm)

- the upward velocity factor $f_v$ for assumed bubble diameters of 3 and 3.5 mm. For comparison, table 6 contains also the upper parameters calculated for the conservative guide values ($p_{1,c}$) of 3 and 4%/m at a diffusor depth of 4.0 m and a supersaturation factor of 1.15.

For the ease of further discussion the above product is named "effectivity number" with the notation

$$E = f_v . d_B^{1,5} (\text{mm}) \qquad \text{mm}^{1.5} \tag{18}$$

A large E-number is obtained by either a large upward velocity (factor) and/or a large bubble diameter. Both conditions hold for a poor aeration system. Large E-numbers, therefore, indicate an inferior, small E-numbers a superior aeration system. The E-number (in contrast to other transfer parameters) allows to compare the most diverse systems of air diffusion by only one parameter. Diversity of the following parameters is included in the E-number

- effective bubble size (definition eq. 18);
- upward velocity (factor)(definition eq. 18);
- diffusor depth (eq. 16);
- the influence of diffusor depth on the depletion of oxygen in the diffused air (eq. 13 and 16);
- rate of air flow and its influences on upward velocity (eq. 8) and oxygen depletion (used in deriving eq. 13 and 16);
- supersaturation factor of the aeration system (many equations);
- temperature correction factor.

For effectively comparing the results of table 6, the obtained E-numbers are classified as follows

| | |
|---|---|
| E > 15 | conservative guide values |
| E > 12 | tests 2 and 5 |
| E > 10 | tests 1, 4, 8, and 9 |
| E > 8 | tests 3, 6, 7, and 10 |

Some class differences can be explained directly by differences in the upward velocity (factor). Conventional spiral flow designs have E-values above 15. Both systems with appreciable circular motion (2 and 5) have E-values above 12.

Due to the lack of information on upward flow velocities and bubble size, differences between systems "E > 10" and "E > 8" cannot be explained rigorously. However, some comparative comments can be made. Comparing tests 3 and 4 (same diffusors, same tank), the better performance (smaller E-value) of test 3 is evident from (a) more domes per $m^2$, (b) better prevention of circular motion by more uniform arrangement of the diffusors, (c) lower (60 % of test 4) average diffusor loadings. Similarly, differences of both horizontal flow systems (9 and 10) can be based on the large difference of diffusor loadings. Systems 6, 7, and 8 differ only in the type of diffusor and E-differences must be assigned to bubble size differences.

With respect to the upward velocity factors (in table 6), it should be

Table 6 Product $f_v.d_B^{1.5}$ and Bubble Velocity Factor $f_v$

| Nr. | $f_v.d_B^{1.5}$ | upw.vel.fact.$f_v$ for | |
|---|---|---|---|
| | | $d_B$=3.0mm | $d_B$=3.5mm |
| 1 | 11.42 | 2.20 | 1.74 |
| 2 | 13.81 | 2.66 | 2.11 |
| 3 | 8.96 | 1.72 | 1.37 |
| 4 | 11.30 | 2.17 | 1.72 |
| 5 | 13.03 | 2.51 | 1.99 |
| 6 | 9.58 | 1.84 | 1.46 |
| 7 | 9.94 | 1.91 | 1.52 |
| 8 | 10.59 | 2.04 | 1.62 |
| 9 | 10.84 | 2.09 | 1.66 |
| 10 | 9.75 | 1.88 | 1.49 |
| 3% | 22.70 | 4.37 | 3.47 |
| 4% | 16.69 | 3.21 | 2.55 |

**mentioned, that the effectivity number is (very) sensitive to (small) bubble diameters. If this factor is not determined from velocity measurements, but calculated from bubble diameters, this information (e.g. from diffusor manufacturers) should be very accurate. For the conditions of test 6, $f_v$-values from 1.59 to 1.73 (table 5) have been determined from velocity measurements. Table 6 reveals $f_v$-values for an assumed bubble size of 3 mm between 1.8 and 2.7 and for $d_B$ = 3.5 mm between 1.5 and 2.1. The conservative guide values lead to $f_v$-values from 2.6 to 4.4 under above assumption, somewhat higher than reported elsewhere (8).**

**Measuring or estimating the upward velocity (factor) is of great practical importance. A large factor (e.g. $f_v > 2$) indicates that improvements of oxygen transfer are possible by more uniform arrangement of the diffusors, by changing the diffusor density and the distance from each other, and by changing the tank geometry. A low value (e.g. $f_v = 1.5$) would indicate that such measures would not have very much effect, since a certain upward velocity cannot be avoided, obviously. Information on only oxygen transfer parameters is not sufficient in this context, because it does not include diffusor depth effect and oxygen depletion. Maximum theoretical transfer values can be estimated (see fig. 5) for comparison, in that case.**

**The application of the effectivity number E to the tests has shown that it is a useful parameter to compare very diverse air diffusion systems operated under different conditions by one quantity only. The error of the approach is estimated to be less than 5%. Disadvantages of E are seen in that it is not dimensionless and that it decreases as the performance of the aeration system improves.**

## SUMMARY AND CONCLUSIONS

A number of full scale aeration tests has been performed in the Netherlands on diffused air systems which try to minimize the upward velocity $v_v$ of the water above the diffusors by either uniform arrangement of the diffusors on the tank bottom or inducing horizontal flow by a seperate water propulsion device. The very diverse test conditions and results are reported and evaluated.

The evaluation indicates that conservative spiral flow guide values of 3 to 4% oxygen absorption per m diffusor depth and 1.8 to 2.4 kg $O_2$/kWh oxygenation efficiency for water ($\alpha = 1$) can be improved by these systems to 4 to 7%/m and 3 to 4 kg $O_2$/kWh (based on 5 Wh/m$^3$ air.m diffusor depth for air diffusion and 1 W/m$^3$ for horizontal flow generation.

Upward velocities $v_v$ were measured in one system (0.13 to 0.17 m/s). Together with an assumed constant slip velocity of small (2 to 6 mm) bubbles in water of $v_s$ = 0,23 m/s, this gives bubble rise velocities in the tank ($v_r = v_v + v_s$) of 0.36 to 0.40 m/s and "bubble velocity factors" ($f_v = v_r/v_s$) of 1.6 to 1.7. A model is developed to quantify the effect of $f_v$ on the rate of oxygen transfer, which predicts the measured values by more than 5% accuracy. It includes the effect of bubble size, diffusor depth, oxygen decrease in diffused air during aeration.

Since in the other aeration systems bubble size and upward velocity (factor) have (as usually) not been measured, both are combined in an "effectivity number" $E = f_v . d_B^{1,5}$ which can be computed from transfer data by the model. The E-number allows to differentiate and compare diverse diffused air systems. Conventional spiral flow systems yield E = 16 to 23, whereas the system tests give E = 9 to 14. For an assumed bubble size of 3.5 mm this would mean bubble velocity factors $f_v$ = 2.6 to 3.5 for spiral flow and 1.5 to 2.1 for the tested systems. E-number and velocity factors indicate whether further improvement of the rate of oxygen transfer (reduction of $v_v$ and $d_B$) is possible by design. Design options in this respect are (a) more uniform arrangement of diffusors, (b) higher diffusor density (elements per m$^2$), (c) lower diffusor loadings, (d) change of tank geometry to further reduce spiral movement. Maximum attainable ($f_v = 1$; $v_v = 0$) are calculated as hypothetical goal (fig. 5). Minimum values of 1.5 ($v_v$ = 0.12 m/s) seem possible.

Above improvements increase at the same time the oxygenation efficiency (kg $O_2$/kWh) to the values proposed above. A further significant increase of OE is limited by the low efficiency of diffusing air into water ranging from only 40% (system 1 at 6.8 Wh/m$^3$.m) to 60% (system 6 at 4.5 Wh/m$^3$.m).

## NOTATION

| | | |
|---|---|---|
| A | m$^2$ | interfacial area between bubbles and water |
| a | m$^2$/m$^3$ | A per m$^3$ of water aerated; specific interfacial area |
| $d_s$ | m | diffusor depth = bubble rise |
| $d_B$ | m, (mm) | diameter of spherical bubble |
| E | mm$^{1,5}$ | effectivity number |

| | | |
|---|---|---|
| $f_c$ | - | oversaturation factor |
| $f_o$ | - | oxygen depletion factor |
| $f_T$ | - | temperature factor |
| $f_v$ | - | upward velocity factor |
| $k_o$ | kg $O_2$/h | empirical constant |
| $K_o$ | g $O_2/m^3_w$.h.m | empirical constant |
| $k_1$ | kg $O_2/m^3$ air | empirical constant |
| $K_1$ | g $O_2/m^3$ air.m | empirical constant |
| $K_L$ | m/h | liquid mass transfer coefficient, also "liquid film coefficient" |
| OC | kg $O_2$/h | oxygenation capacity |
| OE | kg $O_2$/kWh | oxygenation efficiency |
| $Q_a$ | $m^3$ air/h | total air flow |
| $q_a$ | $m^3_a/m^3_w$.h | specific air flow ($m^3$ air/$m^3$ water per hour) |
| $q_d$ | $m^3_a$/ h.diff. | specific air flow ($m^3$ air/h per diffusor) |
| p | %/m | percent oxygen absorption per m bubble rise or per m diffusor depth |
| p' | $m^{-1}$ | fraction of oxygen absorbed per m bubble rise or per m diffusor depth |
| | | for subscripts see context |
| SOC | g $O_2/m^3_w$.h.m | specific oxygenation capacity: g $O_2$ transferred per $m^3$ water per h per m diffusor depth or bubble rise |
| V | $m^3$ | volume of water under aeration |
| $v_r$ | m/s | (geometrical) rising velocity of bubble (in tank) |
| $v_s$ | m/s | bubble slip velocity in water |
| $v_v$ | m/s | upward velocity of water above diffusors |
| $\rho$ | kg/$m^3$ | density of water |
| $\sigma$ | N/m | surface tension of water in contact with air |
| $\tau_B$ | s, h | retention time of bubbles in water |

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# BLOWER DESIGN CONSIDERATIONS

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## ABSTRACT

A major design consideration in diffused aeration is the selection of the air blowers. Over the design life of a treatment facility the organic loading to the aeration tanks will vary significantly creating both lower and higher oxygen demands. The aeration system which can efficiently meet this varying air requirement over the design period of the plant will realize to the owner a considerable cost saving.

To properly select an efficient blower system the designer must appreciate not only the organic loading condition but also the performance characteristics of the blower used, its relationship to other blowers when running in parallel and its relationship to the air system.

## INTRODUCTION

For secondary treatment of wastewater, activated sludge is one of the most widely used and effective processes. Its success, however, depends on an adequate and continuous supply of oxygen to the micro-organisms grown and maintained in the system. Diffused aeration is one of the most proven and reliable methods for transfering oxygen into the aeration tanks. This paper addresses blower design considerations specific to the diffused aeration process.

## ORGANIC LOADING

It is necessary to consider in design the range of organic loading conditions that the treatment system will experience; not just at the design condition but at commissioning, during weekend low strength periods, weekday peak periods in other words throughout the plants design life.

If for example, we consider normal sewage, one not diluted by abnormal contributions of infiltration and inflow, we can expect a sewage strength of perhaps 200 mg/L of $BOD_5$. As flow increases throughout the day variations in organic loading are experienced. Figure 1 shows a typical pattern of loadings to the aeration tanks after primary settling.

In this example you will note that the organic loading as measured in kg/day $BOD_5$ varied from 1500 to 7000, a ratio of approximately 1 to 5.

Throughout the design life of the plant the organic loading will normally increase due to population or industrial growth as shown in Figure 2.

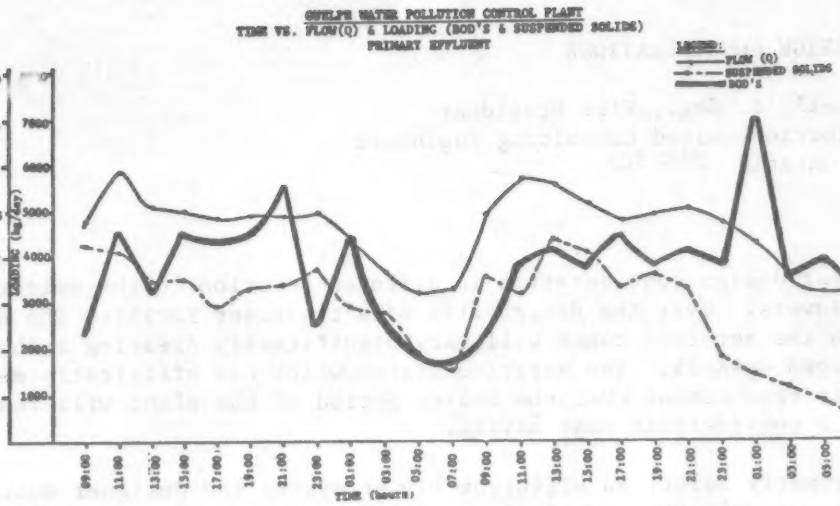


Figure 1. Flow and Load Variations at a Treatment Plant

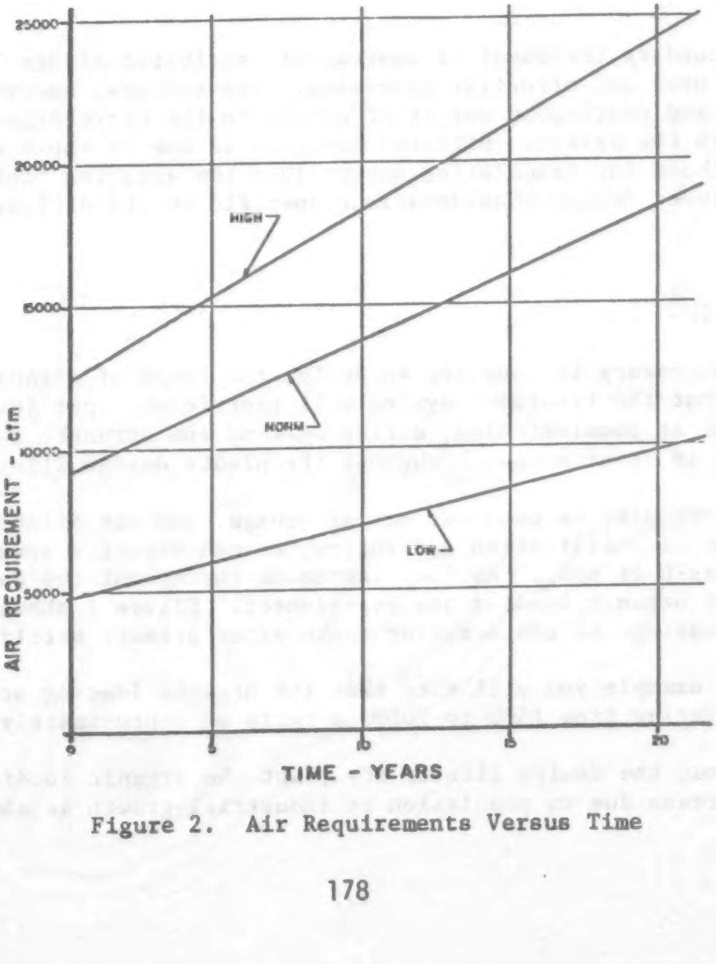


Figure 2. Air Requirements Versus Time

Understanding the range of organic loading condition is therefore an important first step in the blower selection process.

## OXYGEN REQUIREMENTS

The amount of oxygen required to satisfy the aeration tank micro-ogranisms is dependent upon the type and strength of the waste and the activated sludge solids retention time necessary to reduce the waste to the effluent criteria.

Stengel and Shell(1) have proposed the following equation

$$r = \frac{[\ \mathrm{BOD5}\ Q\ ]}{[(1+b)\ \mathrm{SRT}]}\ [A + Ab\ (\mathrm{SRT}) + a\ B\ (\mathrm{SRT})]$$

where $r$ = oxygen required lbs per day
$A$ = oxygen required for BOD (substrate) synthesis, lbs $O_2$/lbs $BOD_5$
$B$ = oxygen required for endogenous respiration, lbs $O_2$/lbs solids
$Q$ = wastewater flow, gal/day
BOD = biochemical oxygen demand at 5 days and 20°C mg/L
SRT = sludge retention time days
$a$ = cell yield co-efficient lbs cells/lbs substrate
$b$ = micro-organisms endogenous decay co-efficient per day

Of note in the above equation are the following:

1. As the activated sludge (substrate) removal rate increases so does the oxygen requirement.
2. As the activated sludge solids retention time (SRT) increases so does the oxygen requirement.
3. As activated sludge synthesis rate yield increases so does the oxygen requirement.

An understanding of the industrial mix of a community is important to this particular item. If there is significant high strength, solubilized wastes expect BOD spikes and higher oxygen requirements.

Treatability studies and knowledge of other industries with similar wastes are most helpful in determining oxygen requirements.

## OXYGEN TRANSFER EFFICIENCY

To arrive at the air supply required the oxygen transfer efficiency of the proposed aeration equipment must be determined. Normally this is done in the laboratory or manufacturers plant by actual oxygen transfer testing under standard controlled conditions. The usual test procedure is the clean water unsteady state method. The efficiency is stated as lb oxygen transferred per hp-hr.

The oxygen transfer efficiency arrived at under the test condition is then converted to field conditions using the following equation:

$$E^1 = \text{alpha } E \, \frac{(\text{Beta Csat} - \text{DO})}{\text{Csat}} \, Q^{CT-20}$$

where $E^1$ = mixed liquor oxygen transfer efficiency
E = clean water oxygen transfer efficiency
alpha = the oxygen transfer correction factor for the particular waste usually 0.75 to 0.95

beta = the correction factor for salinity-surface tension usually 0.95 to 1.00

Csat = oxygen saturation in tank usually taken at mid depth

DO = minimum dissolved oxygen in mixed liquor usually taken as 2 mg/L

Q = temperature correction constant, 1.024

T = temperature of mixed liquor C°

Important factors affecting the rate of oxygen transfer are:

1. Oxygen concentration gradient of the wastewater into and out of the aeration system.
2. Temperature of the wastewater.
3. Turbulence in and around the air/liquid interface.
4. Wastewater characteristics.
5. Depth of wastewater.
6. Contact time of the air bubble and wastewater.
7. Size of air bubble.
8. Rate of air flow.
9. Type of air diffuser.
10. Position of the air diffuser.
11. Aeration tank geometry.

Experience with the actual diffuser device and discussion with a knowledgeable and reputable manufacturer are essential at this stage as shown in Figure 3.

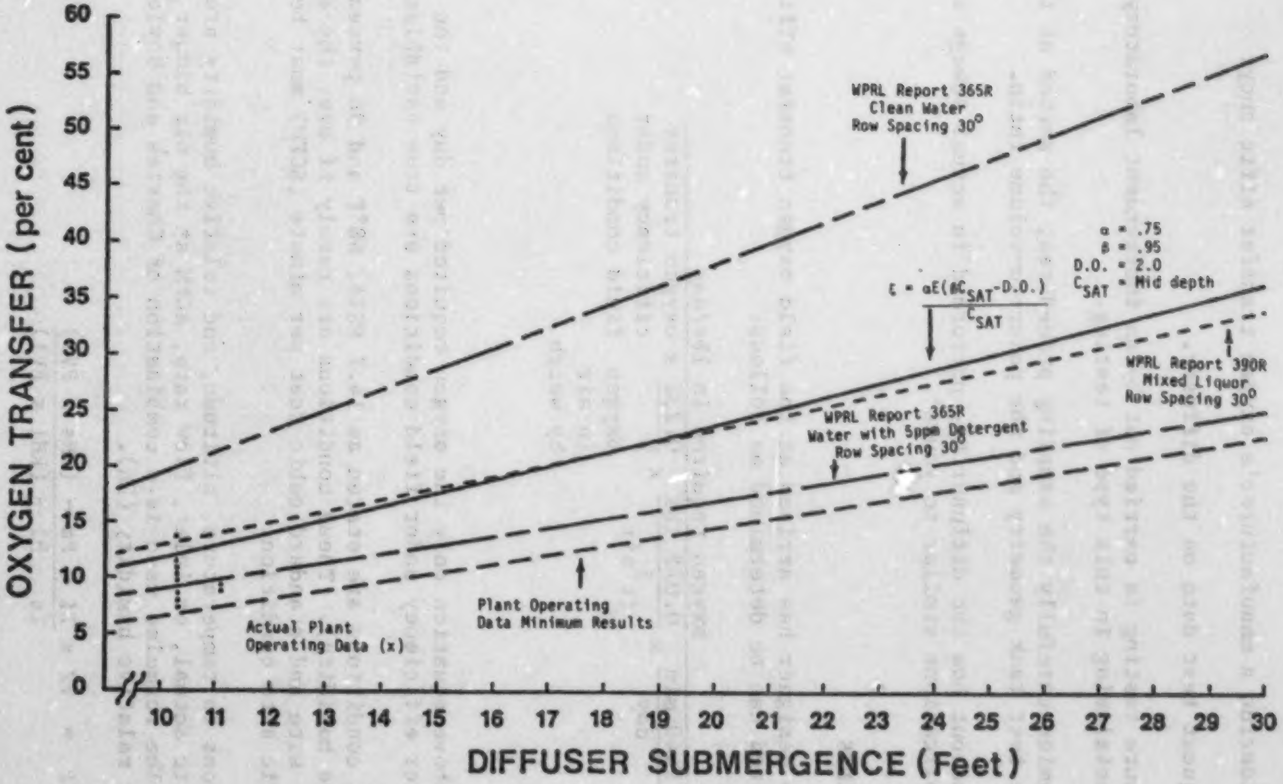


Figure 3. Aeration Device Information

All too often claims of high oxygen transfer efficiencies have not been realized in the field due to the presumptuous claims by aeration equipment manufacturers or designer naivety.

In considering a manufacturer's oxygen transfer efficiency:

1. Request test data on the diffuser.
2. Ensure testing is carried out by an independent laboratory specializing in this type of testing.
3. Examine carefully the sampling procedures, the method of testing, the test tank geometry and the power-to-volume ratio.
4. Find out how the diffuser has performed in actual sewage treatment applications similar to yours'.

## AIR REQUIREMENTS

Once the designer has arrived at the field oxygen transfer efficiency the air required can be determined as follows:

$$\text{Standard Cubic Feet of Air Per Minute} = \frac{\text{oxygen required in lbs/day}}{\frac{1440\ \text{min}}{\text{day}} \times \frac{0.075\ \text{lbs}}{\text{ft}^3\ \text{air}} \times \frac{0.232}{\%\ \text{oxygen in air by weigh}} \times \text{oxygen transfer efficiency under field conditions}}$$

In the above equation only the oxygen required per day and the diffuser oxygen transfer efficiency under field conditions are true variables.

Standard conditions are stated as 14.7 PSIA, 68°F and 36 percent R.H. (relative humidity). These conditions are rarely if ever the actual conditions at site and standard cubic feet per minute (SCFM) must be changed to the specific site condition.

Corrections to temperature, altitude, and relative humidity are made to convert SCFM to actual, or inlet, flow rate, ACFM at the air blower inlet conditions. The formulae used is a combination of Charle's and Boyle's law corrected for relative humidity (RH).

$$V1 = V2 \times \frac{T1}{Ts} \times \frac{Ps - (RHs \times PVs)}{P1 - (RH1 \times PV1)}$$

The above formulae can be simplified to:

$$ACFM = SCFM \times 0.0276 \frac{T1}{P1 - (RH1 \times PV1)}$$

where $V_1$ = actual or inlet cubic feet per minute ACFM at blower inlet

$V_2$ = standard cubic feet per minute at blower inlet SCFM

$T_1$ = Blower inlet temperature in degrees Rankine

Ts = Standard temperature in deg. R. (68°F + 460 = 528°R)

$P_1$ = Blower inlet pressure in psia

Ps = Standard pressure in psia (14.7)

RH = Relative humidity at blower inlet in percent

RHs = Standard relative humidity in percent (.36)

$PV_1$ = Vapour pressure of water at T in psia

PVs = Vapour pressure of water at Ts in psia

## POWER REQUIREMENTS

To determine power requirements the following formulae applies:

$$\text{Brake horsepower} = \frac{w\ R\ T1}{33{,}000\ n\ e}\left[\frac{P2^{n}-1}{P1}\right]$$

where w = weight flow of air lb/min (acfm x .075 lbs/$\text{ft}^3$ air)
R = gas constant (53.5)
$T_1$ = inlet temperature °R
$P_1$ = inlet pressure, psia
$P_2$ = discharge pressure, psia
n = (k-1)/k = 0.283 for air
k = 1.395 for air, specific heat at constant pressure divided by specific heat at constant volume
e = efficiency usual range for compressor is 70 to 80

In order to better appreciate power requirements the cost of power should be based on utility energy escalation projections. In Ontario one horsepower has a projected present worth of $3,500 (Cdn). For a 1000 HP blower one percent difference in overall isentropic efficiency is worth $35,000 (Cdn).

Generally when making blower selection first costs are not the determining factor, but rather, operating costs.

## AIR BLOWER TYPES

There are two types of air blowers commonly used for aeration of sewage:

- Centrifugal
- Rotary Positive Displacement

Selection of which of the two types of blowers to use is generally based on efficiency and cost, however, with increased attention being paid to noise control care must be taken to reduce blower noise levels by careful design and selection of building and air piping materials, blower silencers, etc.

The sound generated by a blower unit is created by all the rotating elements in the blower, motor and speed increaser. The sound is transmitted via the blower casing and foundation, piping system and the motor windage.

Reduction of noise is a major design consideration and can in most cases be reduced to acceptable levels by:

- impeller design
- smooth blower air passages
- blower casing material i.e. cast iron
- dynamic balancing of rotating components
- alignment of blower and motor
- isolation of piping and material
- good foundation design
- isolation of building structure
- gear increaser selection
- silencing on blower inlet and discharge
- sound insulation of air piping
- in acoustical treatment of blower room

The blower size and type should fit into what is generally commercially available. In capacities from 10 CFM to 15,000 CFM. Positive displacement blowers are generally more efficient. Above 15,000 CFM centrifugal machines are almost always used. The reason for this is that centrifugal blowers have higher efficiency at larger volumes of air than do positive displacement blowers.

Where unit capacities are lower than 1,000 CFM positive displacement blowers are preferred and above 10,000 CFM centrifugal blowers are used.

Certain characteristics of the centrifugal and positive displacement blowers are worth noting and are discussed in the following.

## CENTRIFUGAL BLOWERS

A centrifugal blower develops pressure by means of bladed rotating impellers. Air enters the impeller at its eye in an axial direction. The flow is then changed to a radial direction inside the impeller and accelerated toward the peripheral area of the blower casing. Figures 4, 5 and 6 show single stage machines. In a multistage blower air is returned to the eye of the next impeller thus incrementally increasing the pressure. Figure 7 shows a section through a multistage machine. In sewage treatment centrifugal blowers of single stage and up to 3 (multi) stage impellers are employed.

In order to produce pressure in the range of 7.5 psig single stage blowers must turn at speed in excess of 3600 RPM requiring speed increases. Multi-stage machines generally do not have to turn at not more than 3600 RPM and can in a well designed machine run with less vibration and noise than a single stage machine.

One of the benefits of centrifugal blowers is that they are variable capacity, constant pressure machines. Change in capacity is achieved by throttling at the suction or with one particular manufacturer by varying diffusers strategically placed on the pressure side of the impeller. By the above methods turn down capabilities to approximately 45 percent of the rated capacity are possible with resulting savings in power.

Figure 8 illustrates theoretical and actual centrifugal blower chacteristics. Because of internal losses the actual volume output decreases.

The pressure developed within a centrifugal blower is significantly affected by air temperature and density. The highest inlet temperature (lowest air density) produces the lowest pressure at a given inlet volume Figure 9 illustrates the effect of temperature.

With centrifugal blowers care must be taken to avoid surge conditions and this is the low capacity limit for this type of machine. Figure 10 shows the surge point in a typical blower performance curve.

The pressure developed by a constant speed centrifugal blower varies approximately with the inlet air temperature and density. The blower will seek equilibrium between air flow and density. Figure 11 shows resulting changes in the horsepower and pressure-volume characteristics curve. If, for example, the average inlet temperature is 60°F the capacity handled will be at point "b". If the inlet temperature drops to 0°F the air density

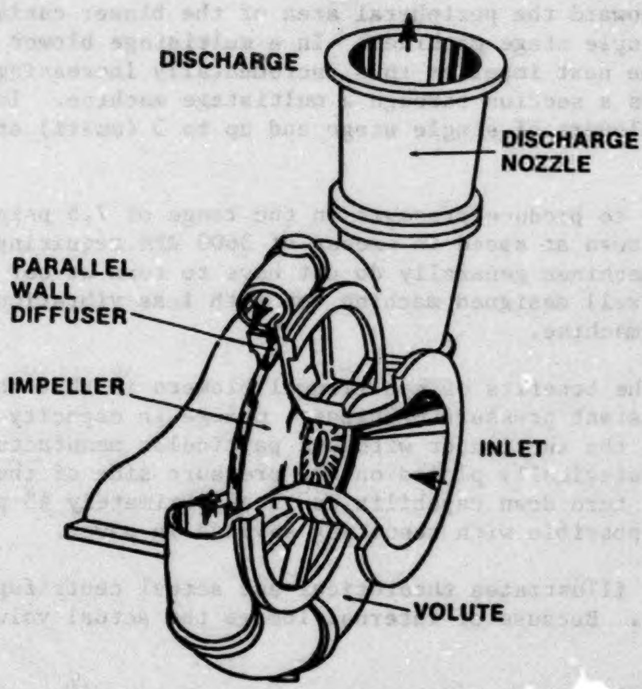

DISCHARGE
DISCHARGE NOZZLE
PARALLEL WALL DIFFUSER
IMPELLER
INLET
VOLUTE

Figure 4. Single Stage Centrifugal Blower

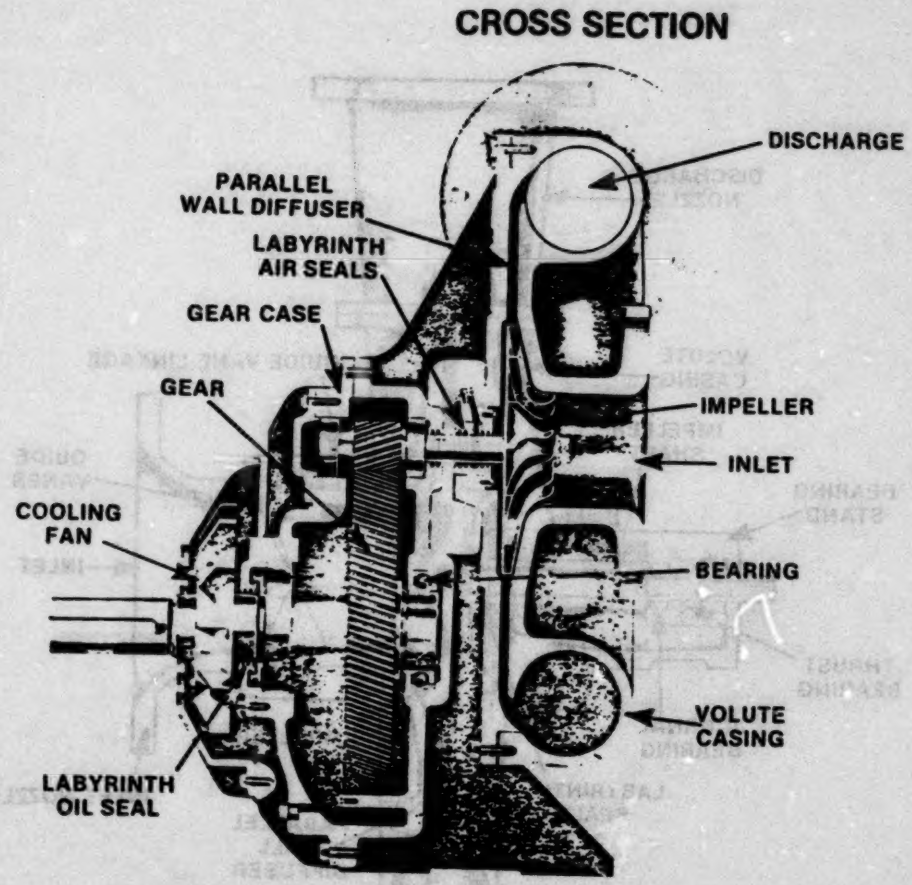

CROSS SECTION
DISCHARGE
PARALLEL WALL DIFFUSER
LABYRINTH AIR SEALS
GEAR CASE
GEAR
IMPELLER
INLET
COOLING FAN
BEARING
VOLUTE CASING
LABYRINTH OIL SEAL

Figure 5. Single Stage Blower with Speed Increaser

**CROSS SECTION**

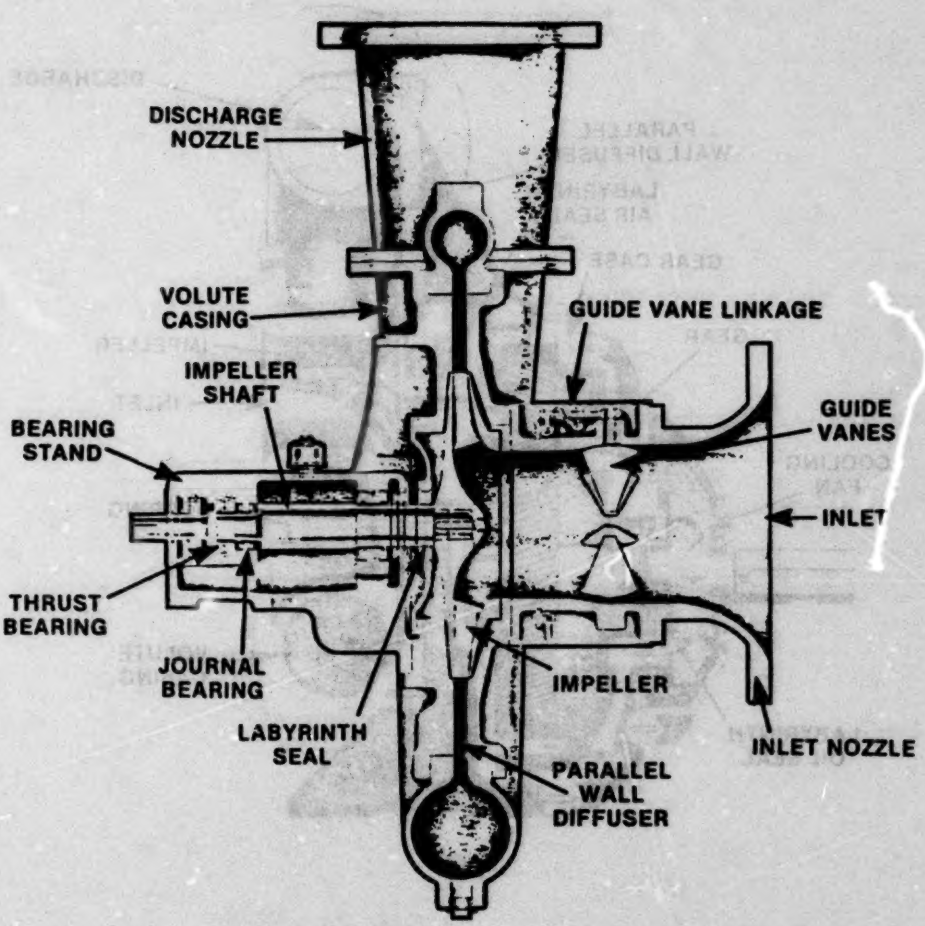
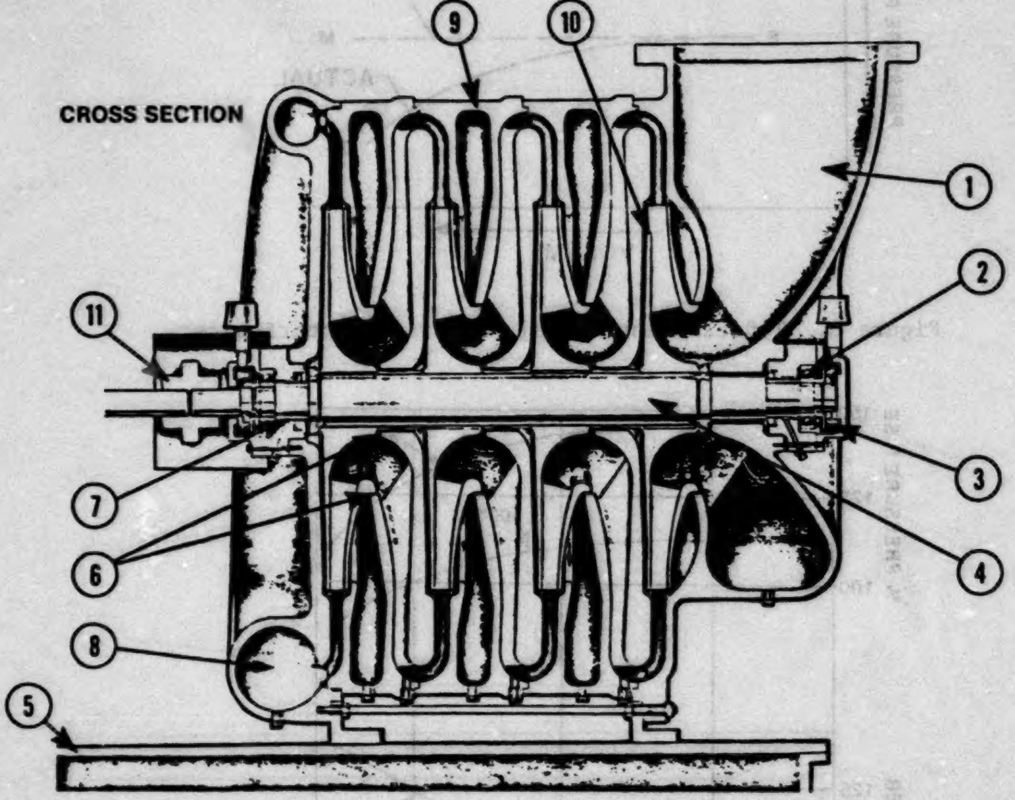

DISCHARGE NOZZLE
VOLUTE CASING
GUIDE VANE LINKAGE
IMPELLER SHAFT
BEARING STAND
GUIDE VANES
INLET
THRUST BEARING
JOURNAL BEARING
IMPELLER
LABYRINTH SEAL
INLET NOZZLE
PARALLEL WALL DIFFUSER

Figure 6. Single Stage Centrifugal Blower with Inlet Throttling



| Part No. | Part Name | Material |
|---|---|---|
| 1. | Inlet Nozzle | Cast Iron |
| * 2. | Anti-Friction Bearing | Steel |
| 3. | Bearing Housing | Cast Iron |
| 4. | Shaft | Carbon Steel |
| 5. | Baseplate | Steel |
| 6. | Labyrinth Seals | Aluminum |
| * 7. | Shaft Seal | Carbon Ring |
| 8. | Discharge Volute | Cast Iron |
| 9. | Case—Diaphragm | Cast Iron |
| 10. | Impellers | Cast Aluminum |
| * 11. | Drive Coupling | Steel |

*Recommended Spare Parts — Bearings, Shaft Seals, Coupling Seals

Figure 7. Multi-Stage Centrifugal Blower

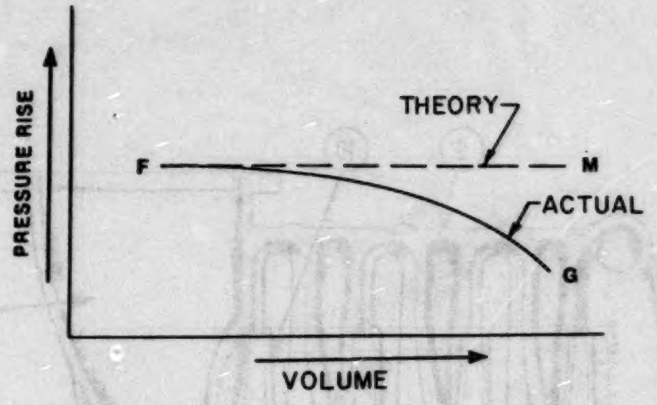


Figure 8. Typical Centrifugal Blower Characteristics

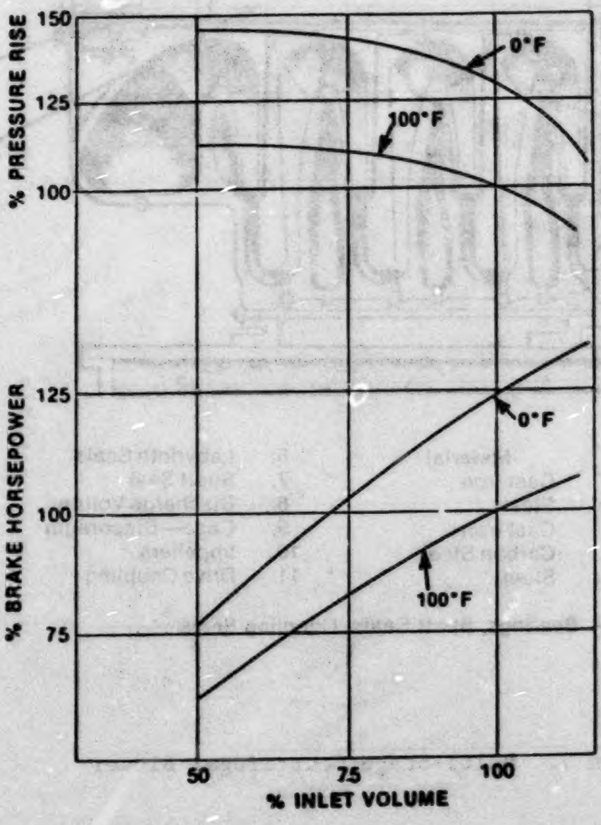


Figure 9. Temperature Effects on Centrifugal Blowers

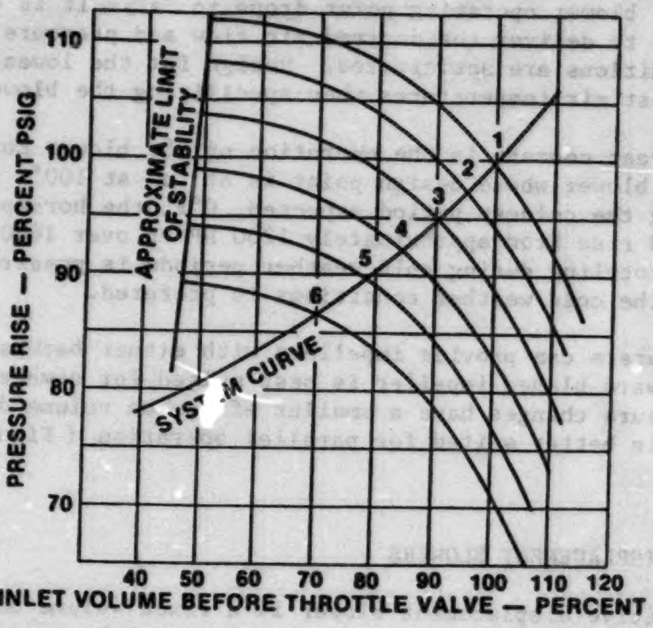


Figure 10. Blower Volume Control By Inlet Throttling

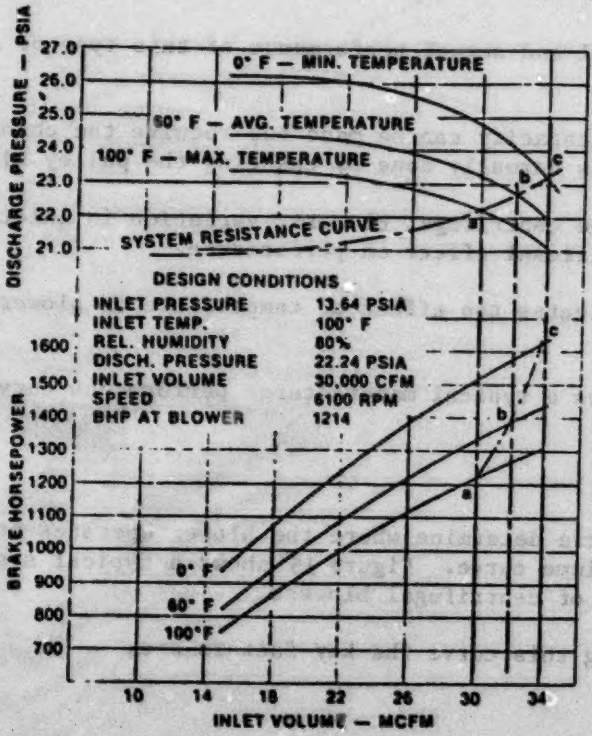


Figure 11. Expected Performance of Single Stage Centrifugal Air Blower

increases and the operating point moves to "c". If the inlet temperature rises to 100°F the blower operating point drops to "a". It is therefore important in order to deliver the desired air flow and pressure that the most adverse inlet conditions are anticipated. Design for the lowest inlet pressure and highest air temperatures when specifiying the blower.

Of just as great concern is the operation of the blower motor. If the inlet volume of a blower whose design point is at "a" at 100°F is left unthrottled during the coldest period expected, 0°F, the horsepower requirements would rise from approximately 1200 HP to over 1600 HP respectively. Throttling during cold weather periods is possible but sizing of the motor for the cold weather conditions is prefered.

The manufacturers can provide impellers with either backward or radial blades. The backward bladed impeller is best suited for sewage treatment operation as pressure changes have a smaller effect on volume delivered and because of this, is better suited for parallel operation ( Figures 12 and 13).

## ROTARY POSITIVE DISPLACEMENT BLOWERS

A rotary positive displacement blower is a fixed volume machine. It develops pressure by trapping a volume of air in two lobe type impellers and carrying the air around the compressor casing to the discharge as shown on Figure 14.

The theoretical and actual performance of this type of a blower is shown in Figure 15.

Variations in capacity can be made but require the change of speed of the blower. This is normally done by changing the pulley size on the blower.

Compared to the centrifugal blowers, variation in inlet air temperatures do not have a significant effect on performance.

Figure 16 indicates the effect of temperature on blower capacity and horsepower.

Figure 17 shows a typical manufactures performance curve.

## SYSTEM CURVE

The system curve determine where the blower operates on a particular blower pressure volume curve. Figure 15 shows a typical system curve with parallel operation of centrifugal blowers.

In calculating this curve the key factors are:

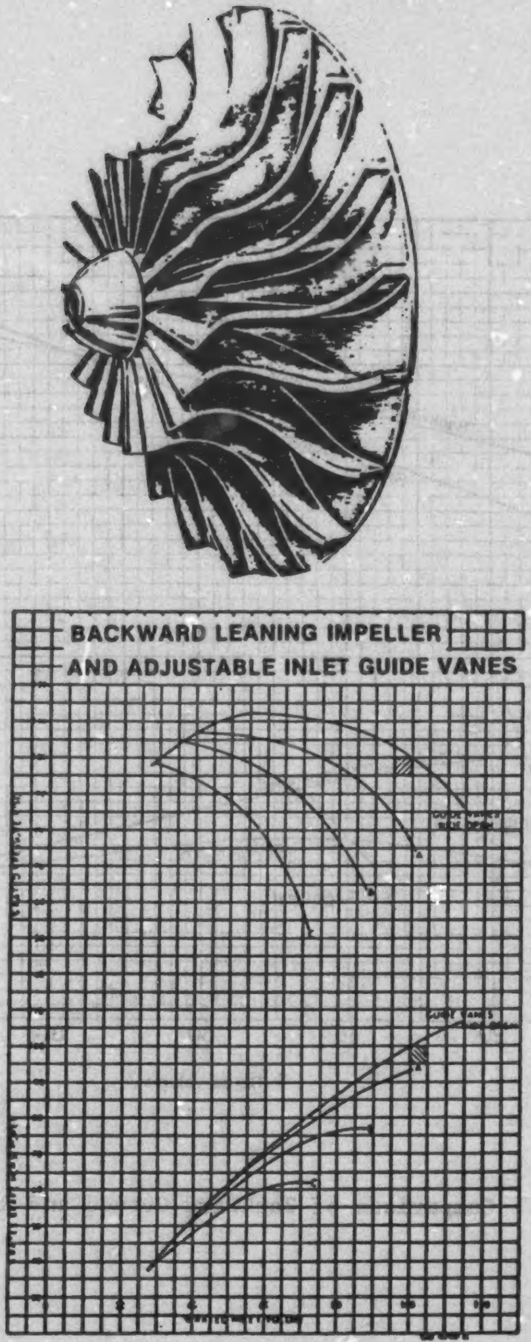

BACKWARD LEANING IMPELLER
AND ADJUSTABLE INLET GUIDE VANES

Figure 12. Open Backward-Bladed Impeller

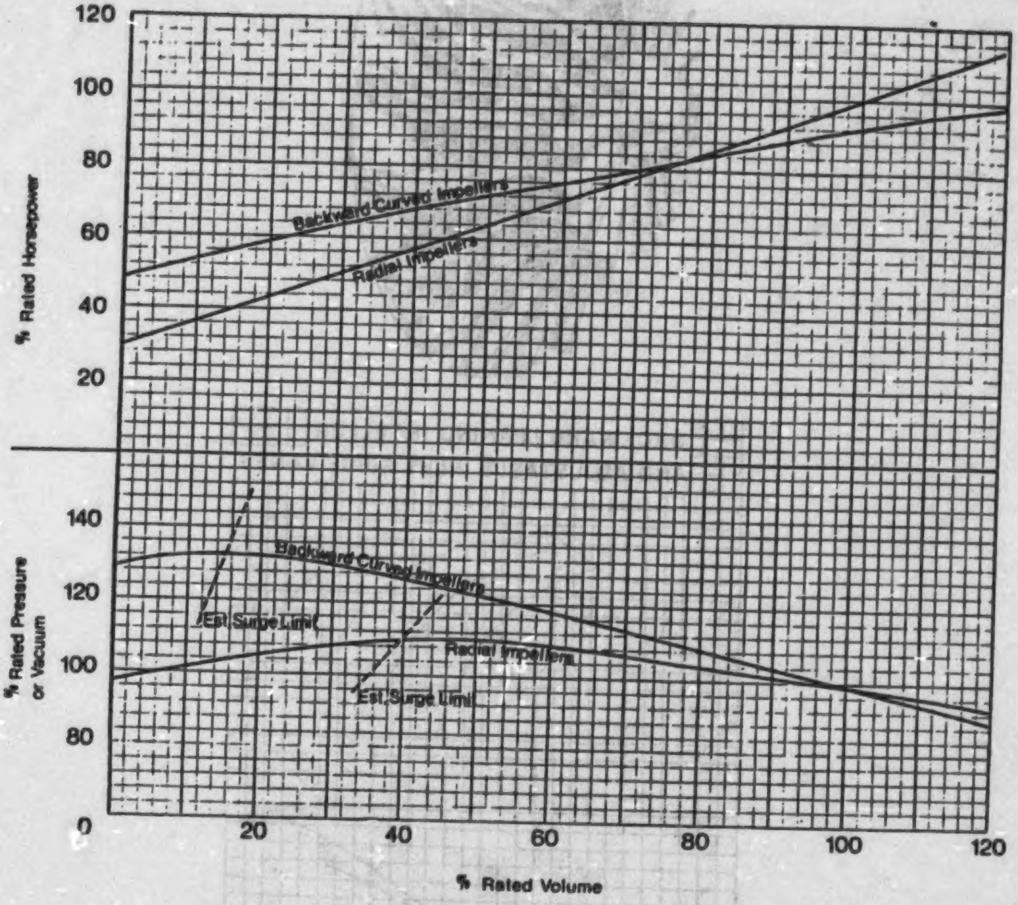
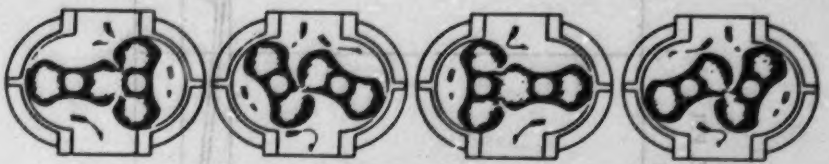


Figure 13. Comparison of Backward Curved and Radial Impellers

OPERATING PRINCIPLE



**Two figure eight lobe impellers mounted on parallel shafts rotate in opposite directions. As each impeller lobe passes the pump inlet, it traps a definite volume of gas and carries it around the case to the pump outlet, where the gas is discharged.**

**Timing gears on each impeller shaft control the relative position of the impellers to each other and maintain the clearances. With constant speed operation, the displaced volume is essentially the same regardless of pressure, temperature or barometric pressure.**

Figure 14. Positive Rotating Displacement Blower

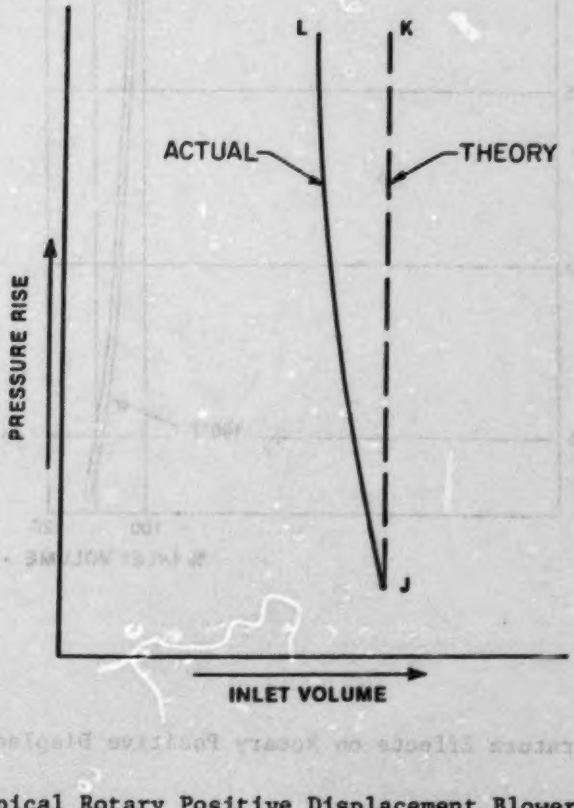


Figure 15. Typical Rotary Positive Displacement Blower Characteristics

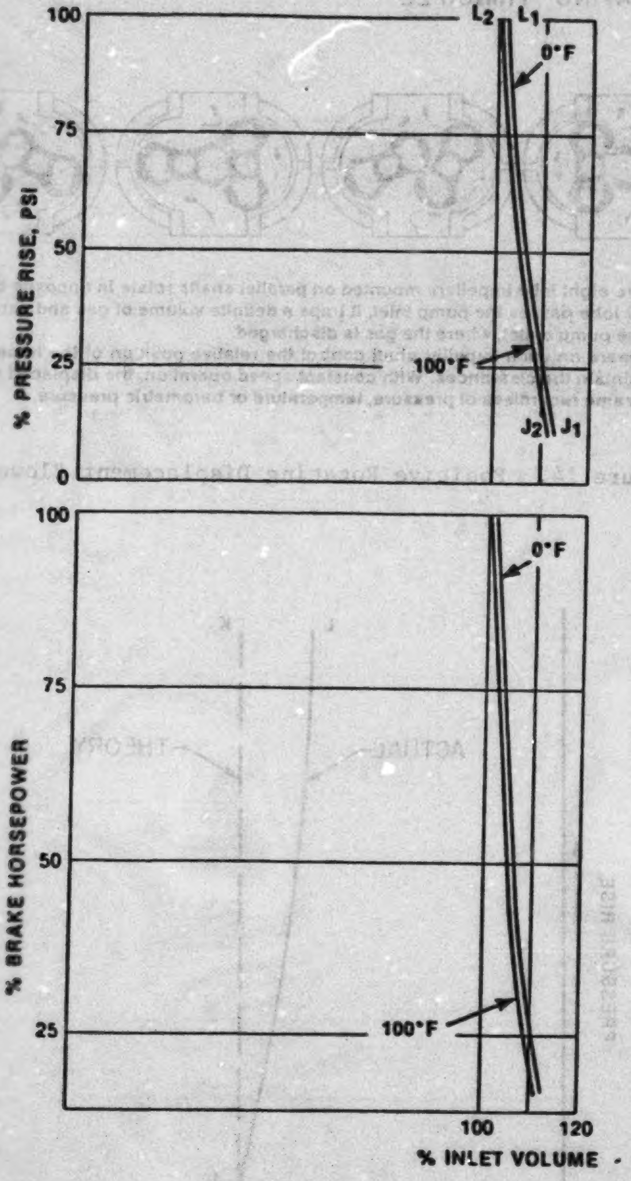
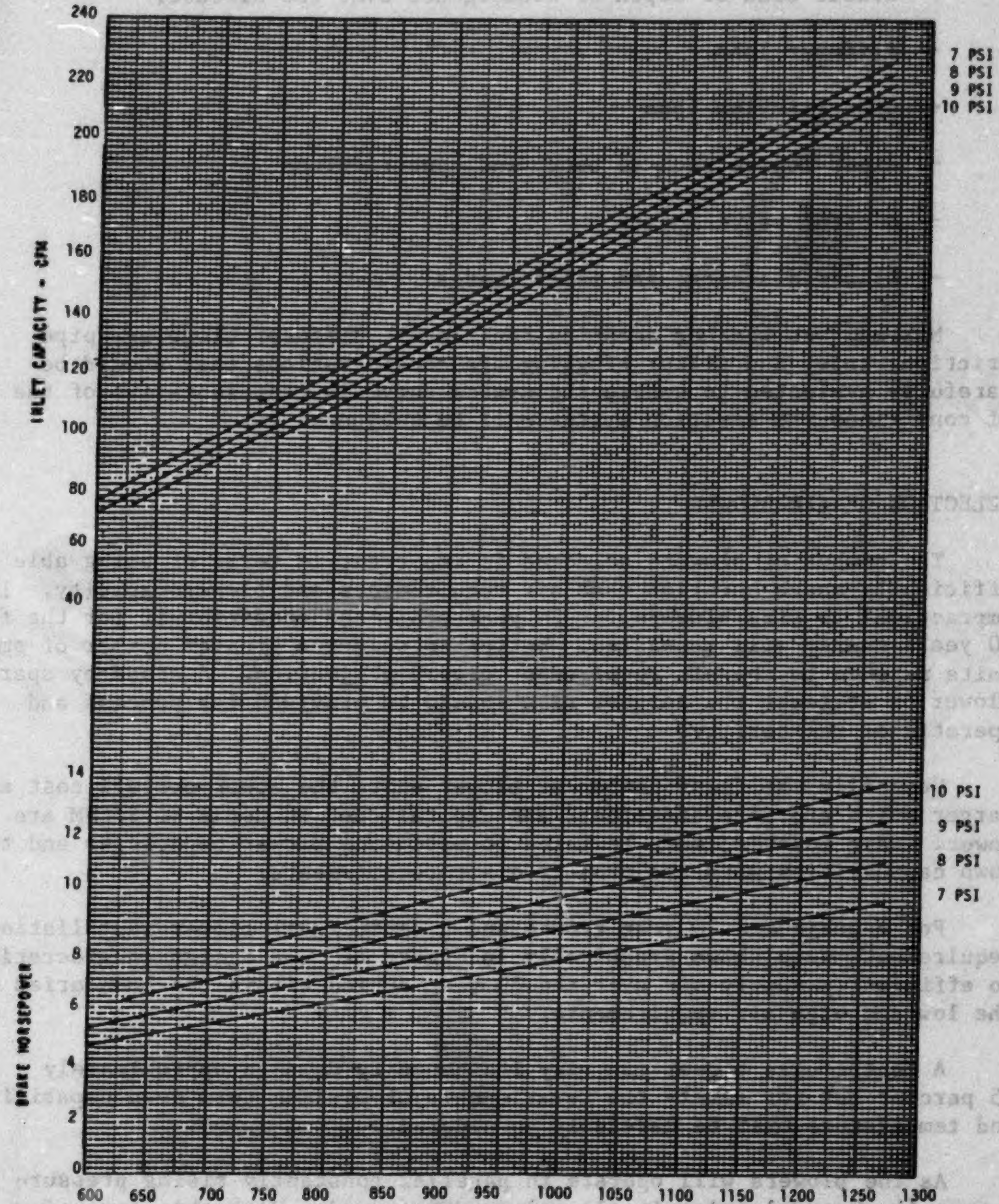


Figure 16. Temperature Effects on Rotary Positive Displacement Blowers



INLET CONDITIONS - AIR • 14.7 PSIA & 70°F

Figure 17. Typical Rotary Positive Displacement Blower Performance

- static head or depth of submergence over the diffuser
- diffuser loss
- pipe friction loss
- minor losses such as bends and constrictions
- control valve losses
- blower discharge and inlet losses

Maximum wet weather aeration tank level, diffuser clogging, pipe friction, inlet and filter clogging are considerations that should be carefully evaluated to ensure the system curve is representative of the range of conditions the aeration system will be subjected to.

## SELECTION OF AIR BLOWER

The number of blowers selected is important in terms of being able to efficiently match aeration tank air requirements and blower capacity. It is impractical to size blowers for 20 year projected condition if for the first 10 years energy will be wasted. Better to select a greater number of smaller units matched to the air requirement growth projection. A stand-by spare blower of at least the largest size should be provided for process and operational reliability.

Generally the fewer number of blower units the lower overall cost as larger units are more efficient, and capital cost in terms of $/CFM are lower. Care however, must be taken to match the blower capacities and turn down capabilities with the projected air requirements.

For a projected 20 year growth and a centrifugal blower installation air requirements were shown graphically in Figure 2. The design consideration is to efficiently supply air over the 20 year period within the boundaries of the low and high air requirements.

A centrifugal blower can vary its capacity down to approximately 45 percent but for safety the relationship of minimum turn down capability and temperature must be carefully considered.

As the blowers will operate in parallel constantly rising pressure curves (backward leaning impellers) are preferred. With this type of impeller slight pressure changes caused by temperature have a smaller effect on the volume of air produced. Figure 18 shows parallel centrifugal blower operation.

Generally it is desireable to have similarly sized machines with one machine as stand-by. For the example chosen the following possibilities arise.

| NO.<br>(includes stand-by) | SIZE<br>CFM | REMARKS |
|---|---|---|
| 2 | 25,200 | Too large can not meet initial low air requirement |
| 3 | 12,600 | Doubtful whether can meet initial low air requirement |
| 4 | 9,200 | Possible |
| 5 | 6,300 | Possible |

Performance curves are considered with particular scrutiny given to parallel operation, turndown capability, efficiency, and temperature.

For the example, the design point is 18,400 CFM and the best fit would be the 9,200 CFM machines. One of the four machines unthrottled would handle the initial average air requirement and could be throttled to provide the initial low air requirement. Two machines both of which would require throttling would meet the high air requirement.

Projected 20 year average air requirements could be met by 2 - 9,400 CFM machines with one machine wide open and possibly with inlet guide vanes over rotated to meet the future low air requirement. Over rotation of guide vanes is possible on larger centrifugal machines and can provide slightly greater volume. The projected 20 year high air requirement would require three machines in operation with slight throttling.

Blower evaluation should consider building, and mechanical requirements. Operating costs should consider future power costs. Life-cycle cost comparisons can be used and provides the designer with a mechanism of comparing overall project lifetime costs. By knowing the total overall cost of two or more blower system alternatives the designer is obviously in a better position to make a better blower selection.

## CONCLUSION

Rising energy and maintenance costs have made the designer aware of the importance of air blower selection. The goal of the designer must be to assure long term reliability and cost effective operation.

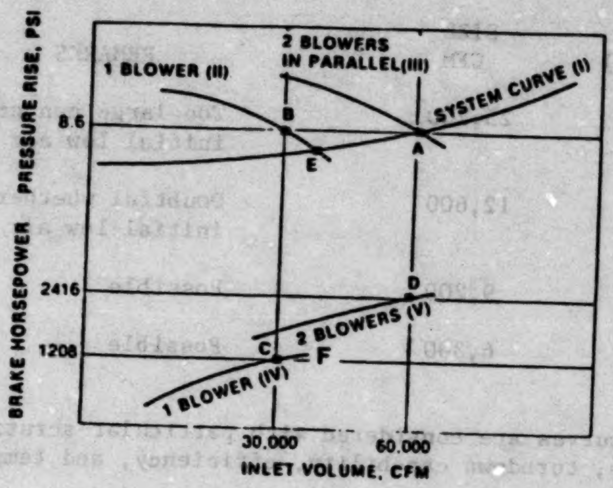


Figure 18. Parallel Centrifugal Blower Operation

## REFERENCE

1. Stensel, H.D. and Shell, G.L. "Two Methods of Biological Treatment Design. Jour. Water Poll. Control Fed., 46, 1, 271 (1974).

## DISCLAIMER

The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency, and no official endorsement should be inferred.

# AUTOMATIC DISSOLVED OXYGEN CONTROL IN THE ACTIVATED SLUDGE PROCESS

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## ABSTRACT

The importance of control over the dissolved oxygen concentration in activated sludge systems is discussed. Dissolved oxygen control is necessary from the viewpoint of energy savings and moreover to improve the monitoring and operational control of these systems. Examples from the literature of these benefits are given.

A brief review of possible control strategy calculations is given. It is shown that the oxygen transfer efficiency may be estimated on-line from dissolved oxygen and power input measurements alone.

## INTRODUCTION

One approach to more efficient operation of wastewater treatment systems is through the application of on-line instrumentation and automatic control. Control, whether automatic or manual, is essential since wastewater treatment systems are rarely, if ever, at steady-state. This is true for the activated sludge process and especially for its aeration component. The need for control of aeration is reflected directly in the cost of energy for this task and in the improvements in process performance and process monitoring that can be accomplished. Clear reference to the advantages to be gained from aeration control has been made by Flanagan and Bracken in the U.S. EPA's design manual concerning procedures and guidelines for the selection of control systems (1). This paper is directed towards an overview of the basic process and computational requirements of dissolved oxygen control in activated sludge systems. For a more detailed review, see the U.S. EPA design manual. Also, an indication is given about the additional process information which can be gleaned from aeration control data by making maximum use of the potential of digital microprocessor capabilities. The proceedings of recent workshops by the IAWPR attest to the great interest in the subject of automatic control for improved operations of wastewater treatment facilities (2, 3).

### Justification for Automatic Aeration Control

The justification for aeration control has been known for some time (4, 5). In 1969, Brouzes (5) indicated that energy consumption in the activated sludge process could be minimized through the automatic control of dissolved oxygen concentration with a spin-off benefit accruing from the automatic control of biomass inventory by relating biomass growth rate to the oxygen transfer rate. Since then, many pilot and full scale examinations of aeration control have been conducted with positive results for energy savings (1, 6, 7, 8, 9). A report financed by the Commission of the European Communities (10) detailed the examination of eight full scale activated sludge systems in Denmark and showed energy savings ranging from 2.5% to 60% through aeration and solids inventory control. The 12 plants within the U.S. study (1) showed an average 33% improvement in BOD removed per kWh of energy supplied. It appears on average, that about 25% of energy consumption could be conserved by aeration control in activated sludge systems but that the actual savings would be site specific. A 1977 survey (11) including 30 activated sludge plants in Ontario, Canada, showed that annual utility requirements comprised 21% of the total plant operating costs. Thus, 25% utility saving could be translated to about a 5% saving in the total plant operating costs by tackling only one major source of expense - aeration. Furthermore, several reports have indicated improved sludge settleability in plants having dissolved oxygen control (8, 12, 13). For example, Wells (8) reported that filamentous organisms were minimized in a dissolved oxygen controlled system compared to a parallel manually controlled system, and that only one-half the clarifier capacity was necessary in the controlled case. Palm et al. (12) have provided clear evidence of the need for further investigation of dissolved oxygen control in the light of its apparent important impact on the growth of filamentous organisms in relation to non-filamentous organisms. They give compelling reasons for the need to consider aeration control over-and-above energy savings. Tuntoolavest et al. (13) gave further justification for minimizing excess aeration due to the complex interaction between aeration shear rates and solids clarification.

Flanagan and Bracken (1) have summarized conditions under which dissolved oxygen control is justified. To avoid the pitfalls of an improper design, their conclusions should be reviewed before deciding upon the selection of a control system.

## ACTIVATED SLUDGE AERATION PROCESS INTERACTIONS

It is known that many interacting factors govern the design and construction of aeration systems, particularly from a steady-state viewpoint. These factors include (14):

- Mechanical design of the aeration system
- Hydraulic disturbances of the influent wastewater and return sludge
- Recycle from other process streams
- Changes in the feed pattern
- Organic load disturbances and substrate storage
- Nitrification/denitrification
- Mixing
- Changes in excess sludge wasting practises
- Disturbances in the oxygen transfer efficiency

It is less appreciated that these interacting factors also govern the effectiveness of an aeration control system in a dynamic rather than steady-state sense. To a certain degree each factor dictates the controller's success or failure. Houck and Boon (15) have documented many examples of the effect of these interactions from a steady-state viewpoint and the problems that can result from inattention to them, especially for plants having fine bubble aeration. Similar examples would apply to other aeration designs.

Flanagan and Bracken (1) provided several examples of design flaws which prohibited the use of installed aeration control systems due to inattention to the complexities of these interactions in the design stage. A typical problem is the oversizing of blowers or air flow control valves to accommodate future plant loads at the expense of current operation efficiency. Improperly dimensioned equipment can result in uncontrollable systems which will not respond effectively to control actions.

Olsson and Andrews (16, 17) have effectively analyzed the interaction between process variables and aeration dynamics to gain insight into the performance and control of the activated sludge process. The relationships between these variables were examined through monitoring only the dissolved oxygen concentration profile and oxygen transfer rate within the aeration tank. Further efforts in the area of improving operational monitoring and control have been discussed by Holmberg (18), Olsson & Hanson (19) and Cook & Libelli (20). In all cases the need for effective aeration control and recursive updating of the oxygen transfer coefficient was established. In these cases, effective use of digital controller capabilities must be made. Of special interest to this workshop is the possibility for the on-line estimation of the process oxygen transfer coefficient, $\alpha K_La$, using the interactions and relationships between the process variables to advantage.

It also is important to differentiate between the long term average steady-state demand for oxygen transfer and the actual non steady state demand due to the plant loads and disturbances. Clifft & Andrews (21) and Sørensen (6) have shown that short term variations in plant organic load are attenuated by the storage and later oxidation of substrate by the biomass. This has important implications for the application of process control strategies, since the magnitude of measured variations in carbonaceous and nitrogenous loads is not immediately reflected in the plant oxygen demand. Thus, the control system hardware should not be designed directly on the measured non steady state carbonaceous and nitrogenous loading and removal, but on providing sufficient oxygen transfer by manipulation of the oxygen transfer coefficient to meet the actual in-situ oxygen demand. The oxygen transfer coefficient is a direct function of the power input to the aeration system and can be varied by manipulating the power input.

Figure 1 shows an example of the relationship between the actual oxygen transfer rate and the variation in carbonaceous and nitrogenous load expressed in oxygen equivalents for a DO-controlled pilot scale activated sludge system treating municipal wastewater (22). The DO concentration was automatically controlled at a setpoint value of 2.0 $g.m^{-3}$. To maintain the DO concentration at 2.0 $g.m^{-3}$ required a significantly greater amplitude in energy input (air flow rate) than was measured for the actual oxygen transfer rate. Similar observations were reported by Sørensen (6).

The relative amplitude of the energy input to the actual oxygen requirement (transfer) is directly dependent on the relationship between the in-situ oxygen transfer coefficient, $K_La$, and the power input. Figure 2 depicts the general form of a linear relationship between $K_La$ and power input as a function of increasing transfer efficiency. In a diffused air system, the power input is proportional to the air flow rate required to maintain the dissolved

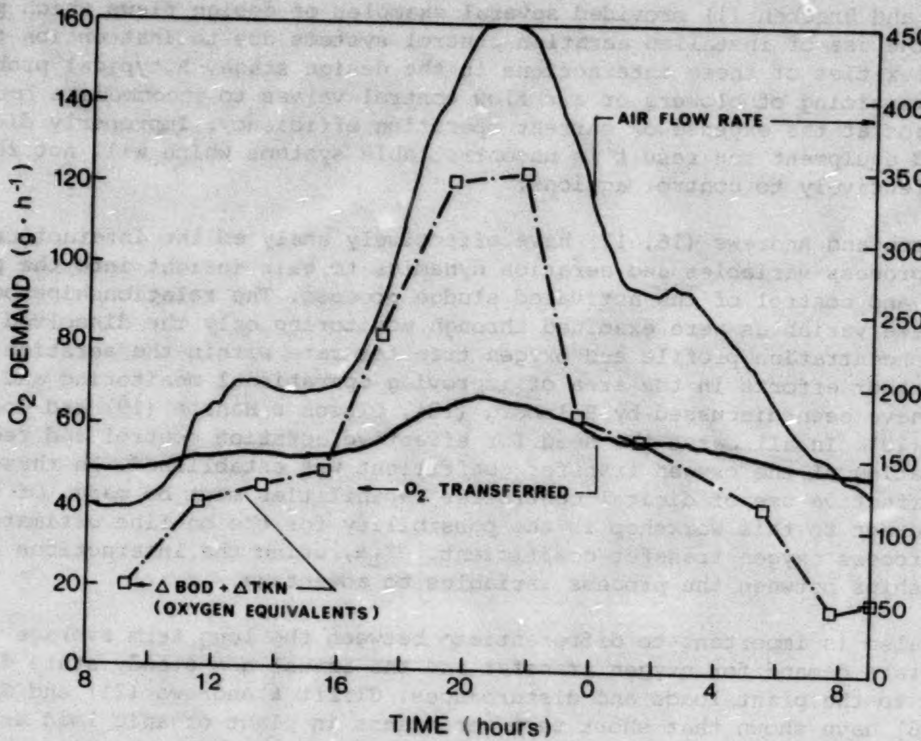

FIGURE 1. Relation between $O_2$ transfer, subtrate removal, and air flow rate under diurnal loading conditions (22).

oxygen concentration. For surface aerators, the power draw depends on the rotational velocity and submergence depth. From the relationship shown (Figure 2), it is clear that the ability to manipulate the oxygen transfer coefficient to control the oxygen transfer rate and the desired DO concentration is entirely dependent upon the ability to manipulate the power input. In the simplest case, the latitude of control achievable is governed by the slope of the relationship shown in Figure 2 and the upper and lower limits on the actual power which can be delivered. Regardless of the system under consideration, it is necessary to be able to mechanically operate the aeration system along an appropriate curve shown in Figure 2 to achieve the degree of control desired under variable loading conditions. It is essential to consider the load characteristics, the primary aeration device, the control elements (e.g. control valves, variable speed drives), and the latitude of control possible to ensure that controllability can be achieved with ease and that real energy savings result. See reference (1) for more detail.

A further consideration in aeration control is the interaction between the aerator and the clarifier (6, 21). Since the respiration of the biomass is the dominant factor governing the need for aeration, the distribution of solids between the aerator and clarifier influences the total oxygen demand in the aeration tank. This is true in a dynamic and in a static sense.

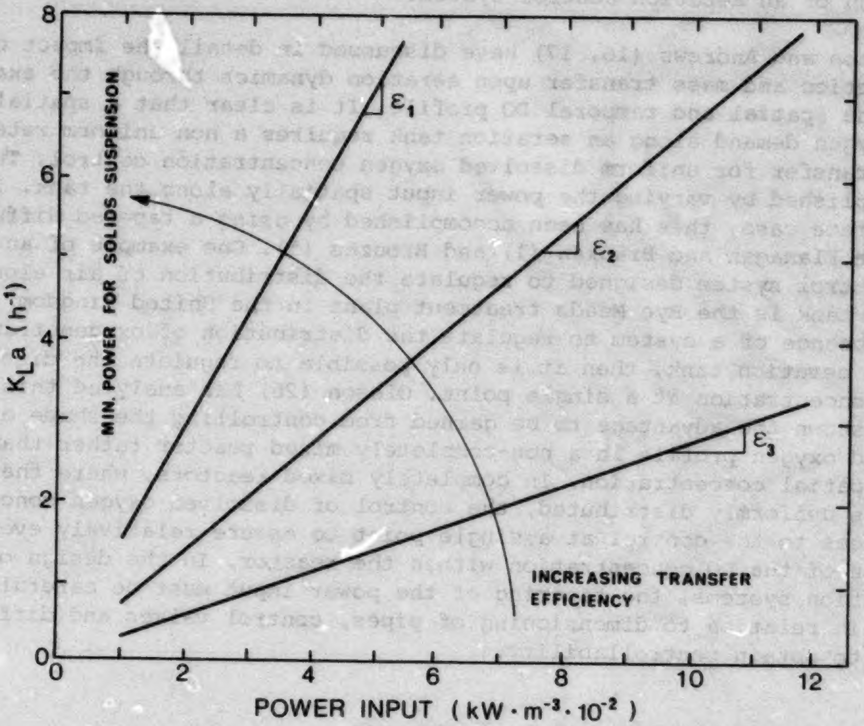

FIGURE 2. Relationship between $O_2$ transfer coefficient and power input for various transfer efficiencies assuming a linear relationship.

If the clarifier capacity to store solids is large and the capacity of return sludge pumping is limited at peak clarifier loading rates, then there will be a transfer of solids from the aerator to the clarifier. This will tend to decrease the oxygenation demand in the aeration tank a corresponding amount. Sørensen (6) has also discussed the impact of clarifier solids inventory buildup upon the total oxygen demand and upon the possibilities for nitrification from a static viewpoint. However, in general there does not appear to be a consensus on the best operating procedure for controlling the distribution of solids between the settler and the aerator. Significant efforts are underway in Sweden to examine rational procedures for solids inventory control (23).

## SIGNIFICANCE OF REACTOR CONFIGURATIONS

A variety of reactor configurations have been developed for the activated sludge process. In general, these designs can be related to the mixing characteristics of the reactors. These characteristics range from a small amount of dispersion to essentially completely mixed (24). Each mixing regime has a significant effect on the aeration dynamics and must be considered in the design of an aeration control system.

Olsson and Andrews (16, 17) have discussed in detail the impact of mixing, reaction and mass transfer upon aeration dynamics through the examination of the spatial and temporal DO profile. It is clear that a spatially varying oxygen demand along an aeration tank requires a non uniform rate of oxygen transfer for uniform dissolved oxygen concentration control. This can be accomplished by varying the power input spatially along the tank. In the steady-state case, this has been accomplished by using a tapered diffuser system, see Flanagan and Bracken (1) and Brouzes (5). One example of an automatic control system designed to regulate the distribution of air along the aeration tank is the Rye Meads treatment plant in the United Kingdom (4, 25). In the absence of a system to regulate the distribution of oxygen transfer along an aeration tank, then it is only possible to regulate the dissolved oxygen concentration at a single point. Olsson (26) has analyzed this situation and shown the advantage to be gained from controlling the shape of the dissolved oxygen profile in a non-completely mixed reactor rather than a single spatial concentration. In completely mixed reactors, where the oxygen demand is uniformly distributed, the control of dissolved oxygen concentration reduces to the control at a single point to ensure relatively even distribution of the DO concentration within the reactor. In the design of overall aeration systems, the tapering of the power input must be carefully considered in relation to dimensioning of pipes, control valves and diffuser density to obtain controllability.

## DISSOLVED OXYGEN/AERATION CONTROL STRATEGIES

In activated sludge process aeration control strategies, the primary objectives are to satisfy the oxygen demand of the respiring biomass and to minimize the consumption of energy. This is almost invariably accomplished by controlling the dissolved oxygen concentration to a prescribed setpoint level. Several studies (1, 4-10, 27) have compared the performance of manual control of the dissolved oxygen concentration to its automatic control, but the conclusions reached have shown that manual control of the oxygen transfer rate is always inferior to some form of automatic control. Manual control leads to about 25% increased energy consumption and widely varying values of the dissolved oxygen concentration. Conversely, effective results in minimizing energy consumption while controlling dissolved oxygen concentration to a setpoint value were reported using closed-loop automatic control to adjust the oxygen transfer efficiency. Stephenson et al. reported a reduction in the standard deviation of the dissolved $O_2$ concentration from 0.9 $mg.L^{-1}$ to 0.2 $mg.L^{-1}$ through DO control at a 2 $mg.L^{-1}$ setpoint (28).

The usual form of automatic control for dissolved oxygen concentration in the activated sludge process has been feedback PI regulation and feedback switching control (on/off or multi-speed) based on the signal from a dissolved oxygen sensor. PI regulation has been satisfactorily used to manipulate variable speed drives (variable frequency type) on positive displacement blowers and on surface aerators, and inlet guide vanes or throttling valves on centrifugal blowers. On/off and multi-speed controllers have been used for intermittently switching drive motors to different set speeds or on and off as necessary. Furthermore, investigations are on-going at the Wastewater Technology Centre and in Sweden (23) to examine continuous self-tuning regulators for the control of dissolved oxygen.

A block diagram representing a simplified feedback control loop for continuous regulation is shown in Figure 3 for dissolved oxygen control. In practice, in the case of diffused aeration with an analog PI regulator, a cascade controller may be required to control the blower output volume and/or pressure to achieve stable and responsive control (1). In such a case the output of the dissolved oxygen sensor is used as a basis for adjusting the setpoint within the inner controller. With a self-tuning controller, it may be that a single feedback loop with dead-time compensation could provide effective aeration control for diffused and surface aeration systems.

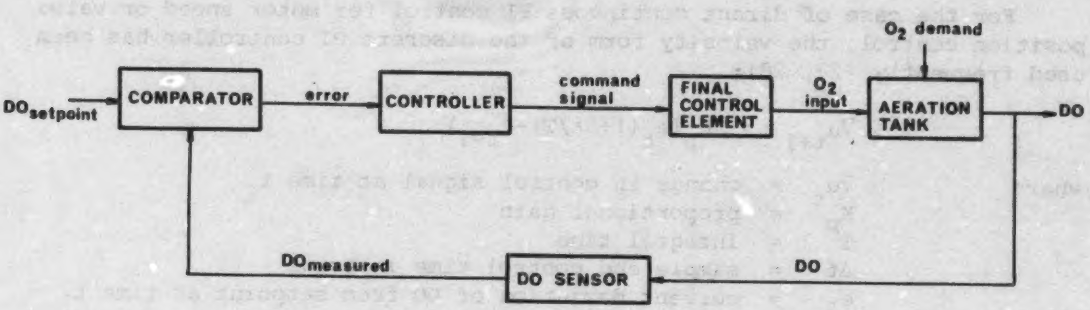

FIGURE 3. Simplified block diagram for DO control system.

Two types of controller hardware, analog and digital, are in use for industrial process control. However, programmable digital controllers are becoming much more popular due to their decreasing price and enormous flexibility. The use of digital controllers offers the opportunity for relatively straightforward dead-time compensation and the implementation of more advanced controllers that can be extended beyond the empirical engineering solutions exemplified by PI control. In addition, raw signals from the process sensors can be digitally filtered to remove undesirable noise. Simple time averaging (7) and exponential filters (19, 27) have been used. The first-order discrete filter resulting from the continuous system differential equation:

$$\tau dc/dt + c = x$$

is used very often in digital systems. Upon discretization this continuous equation becomes:

$$c_{n+1} = \alpha c_n + (1-\alpha)x_n$$

where $\tau$ = continuous system time constant
$c_n$ = discrete filtered signal for n-th time interval
$x_n$ = discrete raw measurement at n-th time interval

$\alpha$ = equivalent filter time constant = $e^{-\Delta t/\tau}$
$\Delta t$ = sample interval

Petersack and Smith (27) used this filter in a demonstration aeration control project with a sample interval, $\Delta t$, of 6 seconds and a value for $\alpha$ of 0.97. This meant that about the last 30 observations $(1/(1-\alpha))$ of the raw DO signal were effectively used in the control algorithm. Wells (8) used combination analog and digital filters to smooth out noise from raw dissolved oxygen concentration measurements. It can be seen that the discrete first-order filter can be programmed easily on a digital controller and needs only two values ($\alpha$ and $c_n$) to be stored between sampling intervals. Filters should be used with care and forethought since a disadvantage of their use is the masking of potentially useful process and system measurement information.

For the case of direct continuous PI control for motor speed or valve position control, the velocity form of the discrete PI controller has been used frequently (27, 28):

$$\nabla u_{t+1} = K_p(e_t(1+\Delta t/T)-e_{t-1})$$

where
$\nabla u_t$ = change in control signal at time t
$K_p$ = proportional gain
$T$ = integral time
$\Delta t$ = sample and control time interval
$e_t$ = current deviation of DO from setpoint at time t.

The actual signal output by the controller is simply calculated:

$$u_{t+1} = u_t + \nabla u_{t+1}$$

Acceptable values for $\Delta t$ in activated sludge systems are between 30 - 300 seconds. Values of the gain and integral time must be obtained from experimental tuning. The velocity form of the discrete controller can be implemented easily on a microprocessor.

Although the PI controller is appealing in its simplicity for automatic regulation, advances in process control have lead to the development of self-tuning minimum variance controllers. These controllers have the general form (29):

$$\nabla^d u_t = - \frac{(\alpha o + \alpha_1 B + \alpha_2 B^2 + \ldots + \alpha_m B^m)}{(\beta o + \beta_1 B + \beta_2 B^2 + \ldots + \beta_l B^l)} y_t$$

where
$u_t$ = control signal
$\nabla^d$ = difference operator, e.g. $\nabla^1 u_t = u_t - u_{t-1}$
$B$ = backward shift operator, e.g. $By_t = y_{t-1}$
$\alpha, \beta$ = controller tuning parameters
$l, m$ = orders of polynomials in $\alpha(B)$, $\beta(B)$ respectively
$y_t$ = observed measurement signal.

Usually d is equal to one and l, m $\leq$ 2. If the correct order of the controller is chosen, which can be established through diagnostic checks, then the controller tuning parameters ($\alpha$, $\beta$) can be calculated and updated recursively as process conditions change, for example due to clogging of diffusers or seasonal temperature fluctuations in aeration systems. Dead-time, also, can readily be accomodated by these controllers. With proper usage, this ensures that the continuous controller remains tuned to track slowly changing process parameters adaptively. These adaptive controllers can be implemented for feedforward-feedback situations and for situations where the sequential changes in the manipulated variable must be constrained (e.g. motor drive changes). This can help not only to avoid process upsets due to rapid and large changes in the control action but also to minimize wear due to excessive control action on mechanical components such as drives and bearings. The usual method of solving the control requirement is through use of a compact recursive least squares algorithm which can be implemented on a digital processor using little space. See Harris (30) for example.

Furthermore, the use of recursive least squares or maximum likelihood algorithms for control or estimation purposes provides an opportunity to estimate process parameters on-line. Under the appropriate conditions, it has been shown that the in-situ value of $K_La$ and volumetric respiration rate (R) can be determined recursively from dissolved oxygen concentration and power input measurements alone (18, 19, 20, 26, 31). For example, Holmberg (18) has examined several ways of estimating $K_La$ for estimation and control purposes in a diffused air system, using the formulation that the oxygen transfer kinetics in a respiring system could be defined as:

$$\tau dy/dt + y = gu$$

where
$y$ = dissolved oxygen concentration
$\tau$ = time constant ($K_La^{-1}$)
$t$ = time
$g$ = system gain parameter
$u$ = air flow rate.

Upon discretization, this continuous system was described by:

$$y_t = \delta y_{t-1} + \omega u_{t-1}$$

where
$\delta = e^{-K_La \cdot \Delta t}$
$\cong 1 - K_La \cdot \Delta t$
$\omega = g(1-\delta)$
$\Delta t$ = estimation time interval
$g = u^{-1}(y_s - R/K_La)$
$y_s$ = saturation dissolved oxygen concentration

Proper selection of the estimation interval, $\Delta t$, and the presence of sufficient disturbance in $u_t$ permitted the recursive estimation of the aeration parameters, $K_La$ and R. In addition to the dynamics of aeration, Olsson and Hanson (19) included a stochastic moving average noise term to account for unknown process disturbances in a recursive model used to estimate $K_La$ at a full-scale activated sludge plant in Sweden. Where the parameters are expected to vary with time (e.g. due to diffusor clogging), then discounted recursive estimation algorithms are used to reduce the impact of old data upon the current parameter estimates. Recursive least squares or maximum likelihood methods can be used. See (19, 29, 30) for example. These methods should track changes in aerator efficiency due to ambient environmental changes. They provide tremendous possibilities for monitoring the actual performance of aeration systems and for improving the control of the activated sludge process (5, 18, 20). Significantly more effort is needed to advance these possibilities. For example, see Yust et al. (32) for the need to estimate respiration rate on-line. To date, no work would appear to have been done to adapt these methods to surface aeration systems.

## AERATION CONTROL IN PRACTICE

The U.S. EPA aeration control manual (1) documented twelve case histories discussing aeration control in the United States in the 1970's. The general conclusions reached indicated that dissolved oxygen control systems were warranted for activated sludge plants having at least 0.044 $m^3.s^{-1}$ hydraulic capacity, if the guidelines presented by the manual were met.

Andersson (10) documented the performance of eight full scale systems in Denmark ranging in hydraulic capacity from 0.0046 to 0.30 $m^3.s^{-1}$ capacity. Except for one of these plants, he indicated economically attractive results through implementing instrument-based feedback aeration control.

For the European and United States' case histories, a variety of aeration devices and control procedures were described. These included diffused and surface aeration devices. Positive displacement and centrifugal blower descriptions have been provided for the case of diffused air, and turbine and brush systems for surface aeration. Control strategies have ranged from continuous PI control of blower output and impeller submergence to on/off control procedures for all aeration equipment types.

Still, it must be said that further development of aeration control policies is necessary. Though strategies exist and have been demonstrated for dissolved oxygen control, their implementation is not widespread in all countries even where the economics may readily warrant it. There are cases where the measurement and control hardware has or will be installed but will only be used for data logging purposes rather than control. Other cases have been known where operators lack even adequate portable instrumentation for monitoring plant dissolved oxygen levels. In these cases maximum aeration capacity has been used at all times even though largely not needed. These examples imply that there are often unnecessary hidden costs not accounted for in plant operation.

There would appear to be the need for more effort toward the control of point source surface aeration systems. Although the hardware to control surface aeration systems does exist, the control strategies for these systems have received less attention than those directed to diffused air systems. In the case of multi or variable speed drives, the designer is faced with the requirement to install these units on several surface aerators compared to a relatively small number of units for blower systems. Efforts are needed to evaluate the design, economy and effectiveness of such systems.

Efforts are underway at the Wastewater Technology Centre in Burlington to implement a feedback self-tuning minimum variance regulator for the control of dissolved oxygen concentration in a pilot scale activated sludge reactor. This will be adapted to a package scale system having a positive displacement blower with variable frequency drive. The results will be used for retrofitting existing larger scale facilities where warranted. Efforts are also underway in Sweden (23) to examine advanced operating control strategies including adaptive control.

In Denmark (33), where aeration control is widely practised, there are studies underway to use the aeration control measurements to regulate the feed distribution in a step feed system using a microprocessor. In this case, the controller will calculate the aeration tank oxygen uptake rate from dissolved oxygen and air flow rate measurements alone. The uptake rate will be used to control feed distribution in an attempt to minimize energy consumption, sludge production and effluent quality.

## SUMMARY

The hardware for aeration control in activated sludge systems exists and should be further exploited to improve plant operations. Although energy savings have been the dominant force behind aeration control, it is becoming recognized that other benefits also are derived from control, e.g. improved settleability.

It also is becoming recognized that other process information can be obtained from aeration control measurements. These include estimation of respiration rate and $K_La$ within the process under time varying conditions. Digital processors must be used to obtain these estimates. With these measurements and tools it should prove to be possible to improve control over such other variables as sludge inventory. More investigation of real-time estimation of process parameters is necessary. Advances in strategies for control of diffused air systems need to be translated to surface aeration systems.

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# EVALUATION OF A DISSOLVED OXYGEN FIELD TEST PROTOCOL

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## ABSTRACT

A test protocol which was developed for on-line dissolved oxygen meters was evaluated in a 60-day field test conducted in an activated sludge aeration basin with thirteen meters from seven cooperating manufacturers. The paper discusses the important features of the test protocol and describes the procedures and results of the field test. These results provide information on methods for testing, calibrating and performance monitoring, information useful for developing specifications, and information on routine maintenance for on-line dissolved oxygen meters in a specific plant environment.

A critical performance requirement for on-line dissolved oxygen meters is to accurately measure dissolved oxygen (within ± 0.5 ppm of $O_2$) with a mean time between maintenance (cleaning, calibration, etc.) of no less than 30 days. Of the meters tested by the developed protocol, nine meters from five manufacturers met the criteria. Of the four meters (two manufacturers) which did not meet the maintenance requirements, one manufacturer has an accessory component which, if made standard, may bring the meter into conformance with the protocol. The other manufacturer has an alternative model which may meet the protocol performance requirements.

The study indicated that the developed protocol is appropriate for acceptance testing of on-line dissolved oxygen meters and that the state-of-the-art meters met the performance requirements of the protocol.

## INTRODUCTION

Continuous monitoring of dissolved oxygen is necessary in wastewater treatment and particularly in activated sludge process control in order to insure satisfactory effluent quality while at the same time conserving energy. Yet only limited systematic information is available on the behavior of commercial dissolved oxygen analyzers in wastewater treatment plant environments. Therefore, the program described in this paper was initiated in order to develop, for a particular set of plant conditions: (1) Information useful for writing procurement specifications for on-line dissolved oxygen analyzers; (2) Methods for accurate and fair acceptance testing, calibration and performance monitoring; and (3) Information on routine maintenance requirements.

The field test was a cooperative effort involving the National Bureau of Standards (Fluid Engineering Division), the Environmental Protection Agency (Municipal Environmental Research Laboratory), the Washington Suburban Sanitary Commission (WSSC), and participating manufacturers of on-line dissolved oxygen meters.

The test document was divided into five sections. Each section provided a detailed description of a specific aspect of the test.

Section one stated the objectives of the test and the test protocol, identified the participants, and defined the responsibilities of each participant during the course of the test.

Section two described the methods used to solicit equipment for the test, and the type of equipment that would and would not be tested.

Section three provided the location of the facility selected for the test, and typical operating parameters for activated sludge process at the facility. This section also stated that each D.O. meter would be assigned a specific location in the aeration basin and that the test period was sixty days.

Section four described the on-site provisions for equipment mounting, installation assistance, and defined the fourteen day period allowed for installation and start up.

Section five described the procedures that would be followed during the sixty day test period, this included the test starting date, the last date manufacturers would be allowed to adjust equipment, the method for determining the level of maintenance that would be provided during the test, the definition of equipment failure, the response to equipment failure, and all information describing the development and logging of test data.

## THE TEST PROTOCOL

A written test protocol was developed to give all participants an opportunity to understand and comment on the proposed test conditions and further to insure that all meter manufacturers would be treated alike. The final test protocol used in this test program reported on here is given in the appendix. It incorporated comments obtained by circulating earlier drafts among prospective participants. Recommendations for additional modifications of the test protocol based on the field tests reported on here are presented in the Conclusions and Recommendations.

### Test Duration

Because many users have indicated that they would consider 30 days a reasonable interval between instrument servicings, a test duration of twice that length, 60 days, was selected as adequate to provide useful information on routine operation and maintenance requirements. To further improve the data base, each manufacturer was required to furnish two meters for the test. A substantially longer test period would have taxed the resources of the staff conducting the tests and would have kept the manufacturers' demonstration units tied up for an inordinate time, while not necessarily furnishing more information on instrument maintenance.

## Responsibilities of Participants

### Manufacturers

To assure proper installation and start-up of all units, the manufacturers were asked to have their own personnel install two identical meters at assigned locations in an aeration basin. Two weeks just prior to the test were allowed for this purpose. These representatives could give written or oral maintenance instructions to NBS and WSSC personnel at the time of installation but were not to be present, unless contacted, once the test period started.

### Washington Suburban Sanitary Commission

WSSC operators were to log the meter readings twice per shift (approximately four-hour intervals) around the clock. WSSC operators were also to perform all "routine" maintenance on the meters (day shift only) as requested and logged by the NBS project supervisor. The definition of routine maintenance was to be determined on a meter-by-meter basis by discussion with the manufacturers at the time of installation. (Membrane/electrolyte replacement was generally agreed upon at that time as the upper bound of maintenance to be provided by operators.) Routine maintenance, such as membrane cleaning, was to be performed as indicated in the user's manual or per oral instructions, but no maintenance requiring the services of instrumentation specialists was to be furnished.

### National Bureau of Standards

NBS staff was to make reference dissolved oxygen measurements at least once a day and be responsible for quality control of the collected data. NBS was also to furnish on-site project supervision with authority to determine, for example, whether meters that failed beyond operator repair should be removed from the test or be serviced by the manufacturer.

## DESCRIPTION OF THE TEST SITE

### The Aeration Tank

The test was conducted at the WSSC Seneca Interim Wastewater Treatment Plant near Gaithersburg, Maryland. This is a 5-mgd plant using five 1-mgd aeration basins operating in extended aeration. There is no primary sedimentation before the activated sludge process and the aeration basins operate with a suspended solids concentration of 4000-5000 mg/liter.

As shown in Figure 1, the aeration basin comprises about seven-eighths of the outer annulus of the 40 m (130 ft) diameter, 5-1/2 m (18 ft) deep tank. The remainder of the annulus is devoted to sludge digestion. The central circular section of the tank is the clarifier, from which samples for probe calibration were taken at the weir overflow. Air diffusers are located radially along the bottom of the annulus at intervals of about 22 degrees. This air flow, which also induces the mixing currents, is kept essentially constant except for

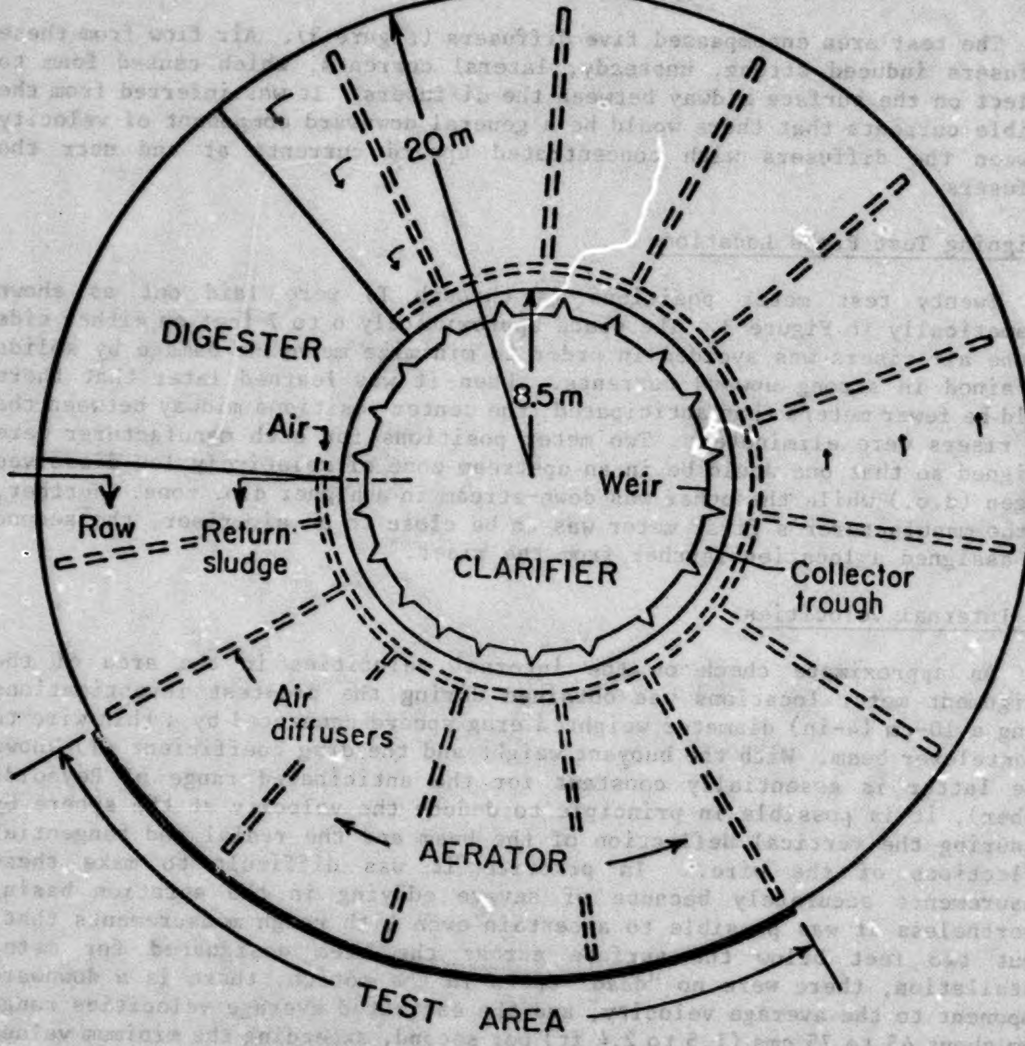
20 m
DIGESTER
8.5 m
Air
Weir
Row
Return sludge
CLARIFIER
Collector trough
Air diffusers
AERATOR
TEST AREA

**Figure 1. Aeration basin test site.**

diversions to the digester section. A test area about 30 m (100 ft) long around the periphery of the tank (Figure 1) was selected mainly because it could be conveniently reached from ground level. An overall view of the test site is shown in the photograph of Figure 2.

The test area encompassed five diffusers (Figure 3). Air flow from these diffusers induced strong, unsteady, lateral currents, which caused foam to collect on the surface midway between the diffusers. It was inferred from the visible currents that there would be a general downward component of velocity between the diffusers with concentrated upward currents at and near the diffusers.

## Assigning Test Probe Locations

Twenty test meter positions (A through T) were laid out as shown schematically in Figure 3. The space approximately 6 to 7 feet on either side of the air risers was avoided in order to minimize membrane damage by solids entrained in strong upward currents. When it was learned later that there would be fewer meters than anticipated, the center positions midway between the air risers were eliminated. Two meter positions for each manufacturer were assigned so that one would be in an upstream zone of relatively low dissolved oxygen (d.o.) while the other was down-stream in a higher d.o. zone. Further, if the manufacturer's first meter was to be close to an air riser, the second was assigned a location farther from the riser.

## The Internal Velocities

An approximate check on the internal velocities in the area of the assignment meter locations was obtained during the pre-test investigations using a 10-cm (4-in) diameter weighted drag sphere connected by a thin wire to a cantelever beam. With the buoyant weight and the drag coefficient (1) known (the latter is essentially constant for the anticipated range of Reynolds number), it is possible in principle to deduce the velocity at the sphere by measuring the vertical deflection of the beam and the radial and tangential deflections of the wire. In practice it was difficult to make these measurements accurately because of severe eddying in the aeration basin. Nevertheless it was possible to ascertain even with rough measurements that, about two feet below the surface across the area designated for meter installation, there were no "dead" spots in the motion, there is a downward component to the average velocity, and the estimated average velocities range from about 45 to 75 cms (1.5 to 2.4 ft) per second, exceeding the minimum values recommended by the manufacturers.

The test area was also surveyed with a portable dissolved oxygen meter as part of the pre-test investigation. No "dead zones" or severe discontinuities were observed along the proposed test locations. However, it was apparent that there were substantial differences in d.o. within the test area (see Figure 4 for later examples) and that individual reference readings would have to be made next to each mounted probe.

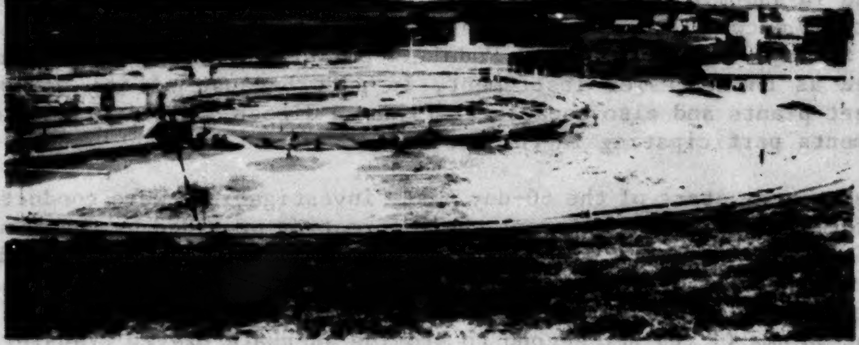

Figure 2. Photograph of test site

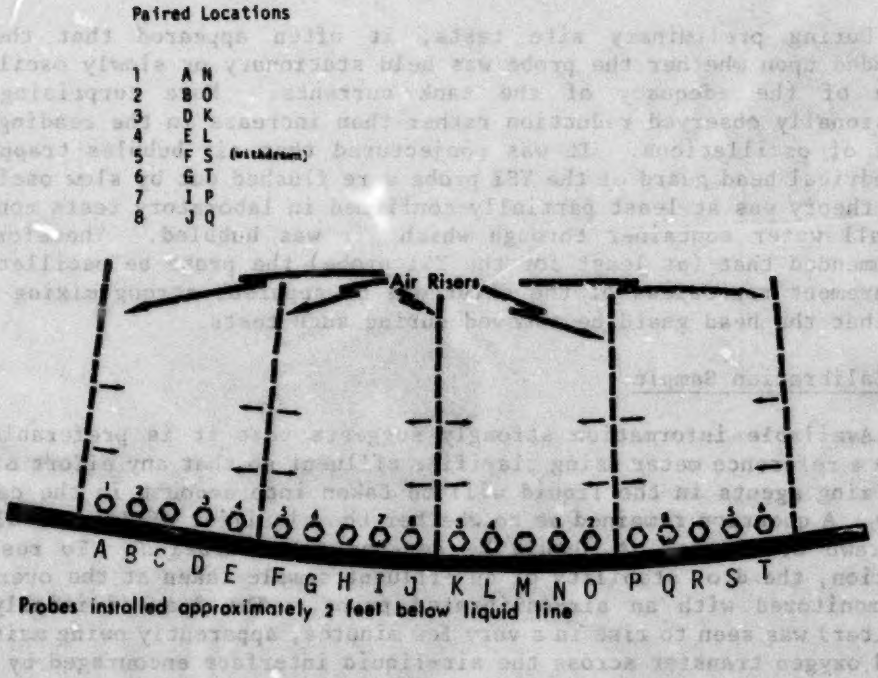
Paired Locations
1 A N
2 B O
3 D K
4 E L
5 F S (withdrawn)
6 G T
7 I P
8 J Q
Air Risers
A B C D E F G H I J K L M N O P Q R S T
Probes installed approximately 2 feet below liquid line

Figure 3. Meter locations.

## REFERENCE MEASUREMENT DEVELOPMENT IN LABORATORY AND FIELD

### The Reference Meter

The use of a portable dissolved oxygen meter as a transfer standard rather than direct Winkler tests was decided on because of the inaccuracies inherent in sampling in the aeration basin and the difficulty in performing accurate Winkler tests directly on mixed liquor samples (2). The combined errors incurred in sampling, sample handling, sample conditioning and analysis would undoubtedly exceed those involved in the careful use of a transfer standard. A Yellow Springs Instrument (YSI) Model 54 portable dissolved oxygen meter was selected as the transfer instrument, in part because of its common use in treatment plants and also because it is not a direct competitor of the on-line instruments participating in the test.

Before the start of the 60-day test, investigations were conducted on the YSI reference meter to supplement the information provided by the manufacturer on estimated measurement errors. The temperature measurement was checked to see that it was within specifications. The user's manual indicated that during measurements the probe should be oscillated axially in the sample at a velocity of about 30 cms (1 ft) per second to provide adequate oxygen transfer. It was noted while experimenting with the probe that the reading would increase by about 0.1 mg per liter if the probe motion was increased to about 60 to 90 cms (2 to 3 ft) per second. During the tests we attempted to calibrate the reference meter using probe velocities close to those in the aeration basin. Nevertheless a potential error source exists here.

During preliminary site tests, it often appeared that the reading depended upon whether the probe was held stationary or slowly oscillated, in spite of the adequacy of the tank currents. More surprising was the occasionally observed reduction rather than increase in the reading upon the start of oscillations. It was conjectured that air bubbles trapped in the cylindrical head guard of the YSI probe were flushed out by slow oscillations. This theory was at least partially confirmed in laboratory tests conducted in a small water container through which air was bubbled. Therefore, it is recommended that (at least for the YSI probe) the probe be oscillated during measurement regardless of the existence of apparent strong mixing currents, and that the head guard be removed during such tests.

### The Calibration Sample

Available information strongly suggests that it is preferable to calibrate a reference meter using clarifier effluent so that any effort of salts or oxidizing agents in the liquid will be taken into account in the calibration (3,4). A question remained as to whether the clarifier effluent could be used as drawn or whether it should be aerated to saturation. To resolve this question, the d.o. stability of an effluent sample taken at the overflow weir was monitored with an air-calibrated probe. The d.o. (initially about 4 mg/liter) was seen to rise in a very few minutes, apparently owing mainly to the rapid oxygen transfer across the air-liquid interface encouraged by the probe motion during the readings. When the test was repeated on an air-saturated sample, the readings were constant for almost one hour. Therefore it was apparent that errors introduced in handling a relatively low d.o. sample to be used for calibration were potentially larger than those introduced (owing to possible non-linearity in the meter) in measuring relatively low d.o. values with an instrument calibrated at a saturation value.

## Winkler Tests

The pre-test period was also used to train the involved NBS staff in the laboratory technique needed for the Winkler (azide modification) test. Many tests were run on distilled water which was aerated with a magnetic stirrer. It was noted that the Winkler result was invariably lower, by amounts up to 3 to 4 percent, than the standard tabular (5) saturation values. Procedures were reviewed and reagents checked and replaced, but the results remained lower; and it was concluded that magnetic stirring for 20 to 40 minutes under the existing ambient conditions did not saturate the sample to its theoretical limit.

## Error Estimates for Reference Readings

In order to compare fairly the readings of the manufacturers' meters with the reference readings, it is necessary to have an estimate of the accuracy of the latter. Potential error sources include errors inherent in the instrument itself, errors due to calibration and drift, and errors unique to this type of test.

Errors inherent in the instrument, stemming mainly from non-linearity and temperature compensation, were estimated from information furnished in the YSI manual to total (quadratically, i.e., square root of sum of squares) about 0.12 mg per liter.

The potential errors in calibration include imprecision in the Winkler average (estimated at 0.05 mg/l) and the previously mentioned effort of velocity on the membrane. The total quadratically summed calibration error is estimated to be not larger than 0.12 mg/liter.

During the field tests any reference meter drift, whatever its cause, was monitored by checking the calibration after the reference measurements were completed. The change rarely exceeded 0.1 mg/liter and usually was smaller.

The error sources specific to this type of test stem from spatial (Figure 4) and temporal variations of dissolved oxygen in the aeration tank. In practice the reference probe had to be a small distance, about 30 cms (1 ft), away from the mounted probe. The resulting uncertainty was assigned an estimated value of 0.05 mg/liter. Also, at a fixed point, dissolved oxygen fluctuations of varying amplitude and period were registered on both the mounted and the reference meter. The magnitude of the resulting inaccuracy depends on the success of the observer in determining a meaningful reading for both meters. To attempt to quantify the error, instantaneous readings at 10-second intervals were made on the reference and mounted meters at several stations. Figure 5 shows one example. Some stations showed smaller variations, but it is clear that a single observer making visual readings over a limited time introduces an uncertainty which was estimated to be about 0.1 mg/liter.

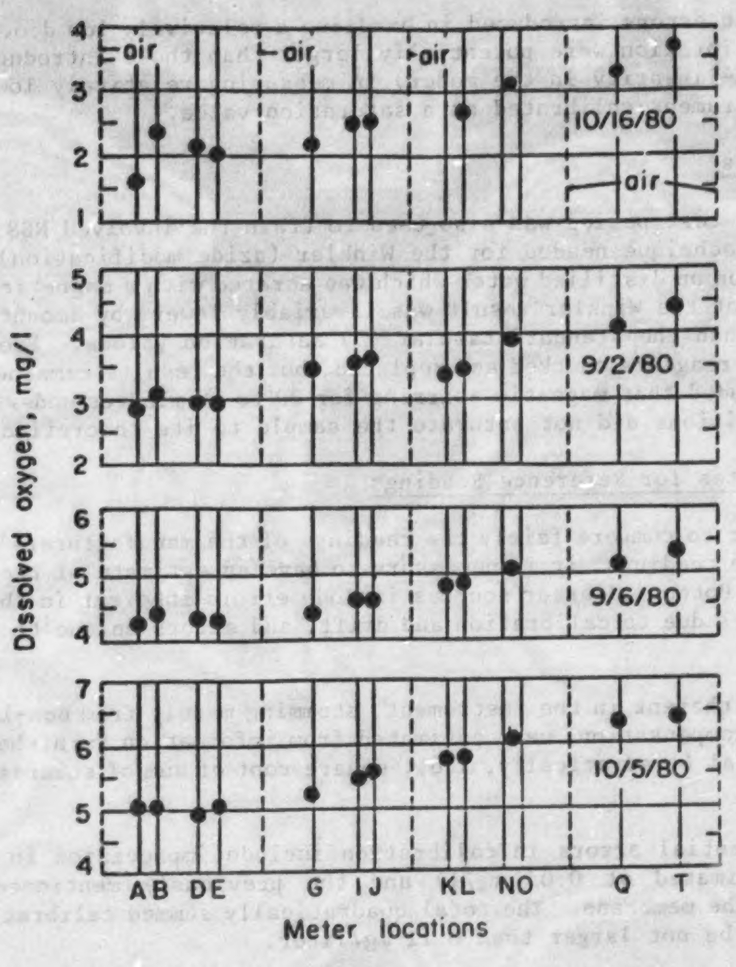

**Figure 4. Examples of reference readings.**

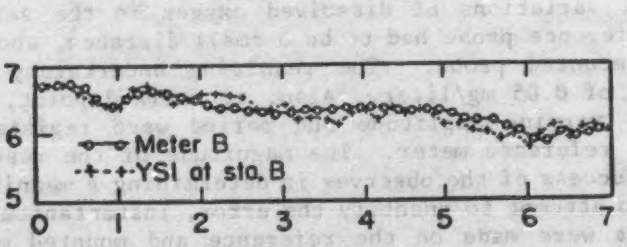

**Figure 5. Example of D.O. fluctuations at a point.**

The errors estimated in this section are summarized in Table 1. The way in which they are combined may vary in specific tests and should be agreed upon at the start. However, the likelihood of all errors being in the same direction appears remote, so that an arithmetic sum may not be realistic. A quadratic summation, which gives a combined error of about 0.23 mg/liter, was used for this test.

TABLE 1. ERROR ESTIMATES FOR THE REFERENCE MEASUREMENTS

| Source | Error, mg/l |
|---|---|
| Instrument | 0.12 |
| Calibration | 0.12 |
| Drift | 0.10 |
| Probe Distance | 0.05 |
| Fluctuations | 0.10 |

## THE FIELD TESTS

### Test Meters

Thirteen meters were installed by seven manufacturers (one did not have a second demonstrator available) in the locations shown in Figure 3. All of the probes used electrochemical cells separated from the sample liquid by a membrane permeable to oxygen. To the extent that descriptions were available in the user manuals, both passive and active cells were represented as well as different electrode materials and electrolytes. Those membranes for which information was given were 2 mils thick, except for one which was 0.1 mm thick. (Membrane thickness will affect response time.) One pair of probes used "permanent" membranes. All of the probes except for one pair were basically cylindrical shapes (with the membrane and cell at one end) to be immersed directly in the ambient liquid flow. The remaining pair used a continuous flow of tap water to induce sample flow through a short pipe which contained the sensor cell in its wall. The participants of the following manufacturers in a test that did not involve a procurement action is sincerely appreciated: Beckman Instruments, Incorporated (Model 7002); Envirotech Corporation, National Sonics Division, Delta Scientific (Series 8210/8310); Fischer & Porter (Series 17D01000); Leeds & Northrup (No. 7931); Orbisphere Laboratories (Model 2716); Rexnord Instrument Products (Model 62 Probe, 3000-5 Analyzer); Uniloc Division of Rosemount, Incorporated (Model 424).

### Procedure for the Daily Tests

A fresh 2000 mm sample of clarified effluent was aerated in a beaker for 20 to 30 minutes with a magnetic stirrer. Three 300-ml B.O.D. bottles were filled with this aerated sample for Winkler tests, and the remainder of the sample was retained for meter calibration and the post-test calibration check.

An effort was made to keep the sample temperature constant. The arithmetic average of the Winkler analyses was used to calibrate the reference meter.

The calibrated meter was carried to the nearby test site and the reference probe was immersed next to the on-line probe mounted at station A. A mixed-liquor temperature reading was made at this point. Care was taken to assure that the reference meter had attained a steady state before meter readings were recorded. This precaution had to be observed at all stations, although the stabilization time was considerably shorter in cases where the reference probe did not have to be removed from the liquid when changing stations. The reference readings at each station were made over a period of about one minute, depending upon their steadiness. The complete set of reference measurements was usually completed within 25 minutes. The reference probe was then rinsed and the meter was carried back to the laboratory, where its calibration was checked in the original aerated effluent sample.

Auxiliary Measurement for Quality Assurance

At random during the test period, the reference meter was carefully air calibrated according to the manufacturer's instructions and its reading in the aerated effluent sample was noted prior to adjusting the calibration setting to conform to the Winkler average. Also, at random a one Winkler test was made on distilled water aerated to saturation, and an air-calibrated probe reading was then taken in the same sample. These tests were designed to monitor the state of the Winkler procedures and reagents as well as the condition of the probe.

Both of the foregoing sets of auxiliary measurements formed a body of data on the efficacy of air calibrations. Examples of the measurements with an air-calibrated probe in the clarifier effluent are shown in Table 2. The results of 28 measurements showed the probe readings exceeding the Winkler values by an average of 4 percent ± 2 percent standard deviation. Examples of the air-calibrated probe measurements in aerated distilled water are given in Table 3. The results of 33 air calibrations on 20 distilled water samples showed the probe readings exceeding the Winkler values by an average of 1 percent ± 1.4 percent standard deviation. It should be noted that these auxiliary tests were somewhat handicapped by lack of a temperature controlled environment.

TABLE 2. EXAMPLES OF AIR-CALIBRATED PROBE VS. WINKLER IN SATURATED CLARIFIER EFFLUENT

| Test Date | Probe (mg/l) | Winkler (mg/l) | Ratio (Pr/Wi) |
|---|---|---|---|
| 10-20-80 | 7.68 | 7.54 | 1.019 |
| 10-14-80 | 8.90 | 8.67 | 1.027 |
| 10-06-80 | 8.52 | 7.85 | 1.085 |
| 9-26-80 | 8.45 | 8.35 | 1.012 |
| 9-22-80 | 8.20 | 7.87 | 1.042 |
| 9-09-80 | 7.12 | 6.64 | 1.072 |

TABLE 3. EXAMPLES OF READINGS IN AERATED DISTILLED WATER

| Test Date | Winkler (mg/l) | Std. Sat.(*) (mg/l) | Winkler Std. Sat. | Probe Std. Sat. | Probe Winkler |
|---|---|---|---|---|---|
| 9-26-80 | 8.64 | 8.78 | .984 | .986 | 1.002 |
| 9-23-80 | 7.59 | 8.00 | .949 | .974 | 1.026 |
| 9-29-80 | 8.77 | 9.04 | .970 | .973 | 1.003 |
| 9-30-80 | 8.74 | 9.07 | .964 | .980 | 1.017 |
| 10-15-80 | 8.96 | 9.18 | .976 | .996 | 1.033 |
|  |  |  |  | .967(**) | 0.991 |
| 8-21-80 | 8.04 | 8.39 | .958 | .985 | 1.009 |
|  |  |  |  | .947 | 0.989 |

(*) From "Standard Methods"
(**) Probe tests repeated when air calibration did not check after reading in distilled water

Typical Test Data

Figure 6 shows the entire 60-day record of operator readings for one meter. Also shown are the two points at which this meter was re-spanned (recalibrated) downward when its deviations from the reference readings began to exceed about twice the estimated reference meter error (0.4 or 0.5 mg/liter). This performance typified that of many of the meters.

Figure 7 shows details of a six-day portion of the record for two meters, B and O, from the same manufacturer but located about 17 m (55 ft) apart. The expected tendency toward a diurnal d.o. variation is visible here as well as the increase in average d.o. at the downstream station. The two sets of values are seen to be in phase and, in this case, about 1.0 mg/l apart.

Figure 8 shows examples of the 60-day record of differences between the daily reference readings and the simultaneous meter readings, positive deviations indicating that the meter value was high. Longer term trends estimated from five-day averages are also shown. The trends are interesting in that occasionally (see K, for example) the deviations from reference values would recede without adjustment after a period of larger excursions. In the case of meter L, some of the excursions are due in part to relatively large oscillations in its readings. The user manual describes a cell modification to reduce these transients.

Maintenance Summary

Each meter was calibrated by the manufacturer's representatives at the time of installation, and the project supervisior was asked to re-span the meters if necessary to agree with the reference reading on the first day of the test. Oral instructions further suggested that the probes need not be routinely removed from the tank unless a need was indicated by poor readings. This policy was followed, except for meter I, for which the membrane was cleaned twice to see if there was any effect. There was none.

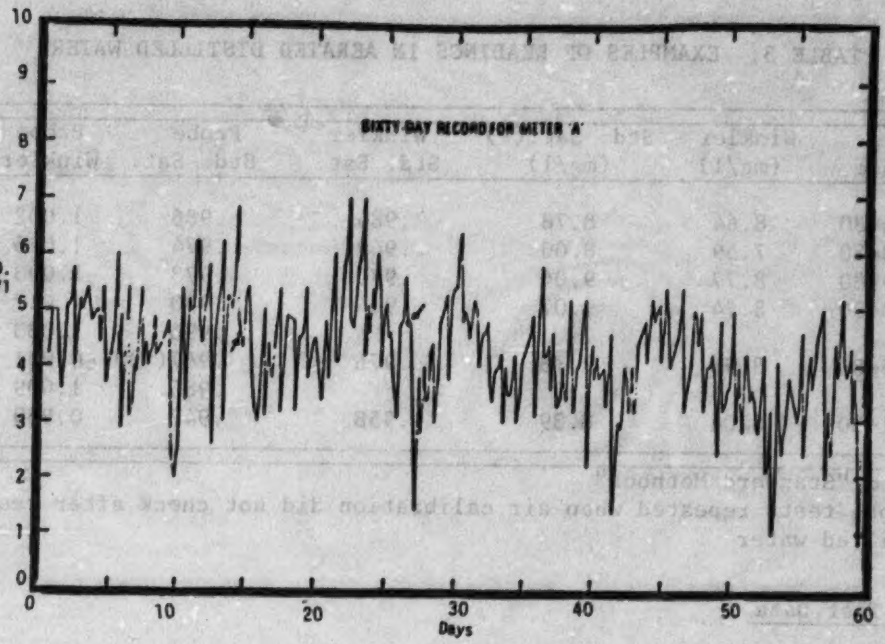

Figure 6. Example of a 60-day meter record.

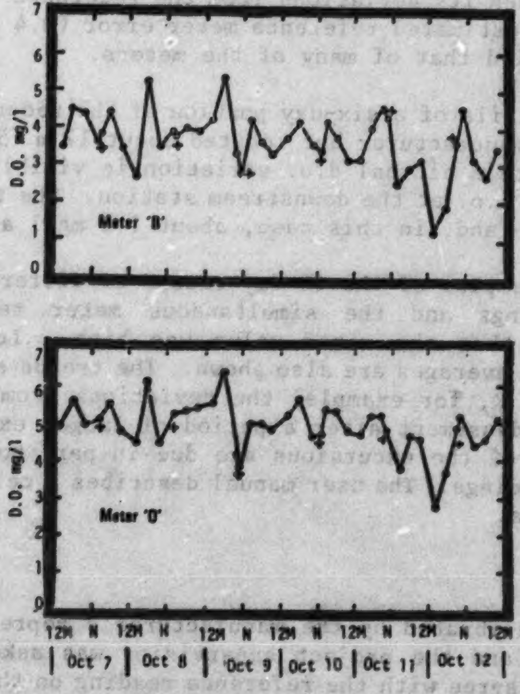

Figure 7. Six-day record for two meters of the same make.
+ = reference reading.

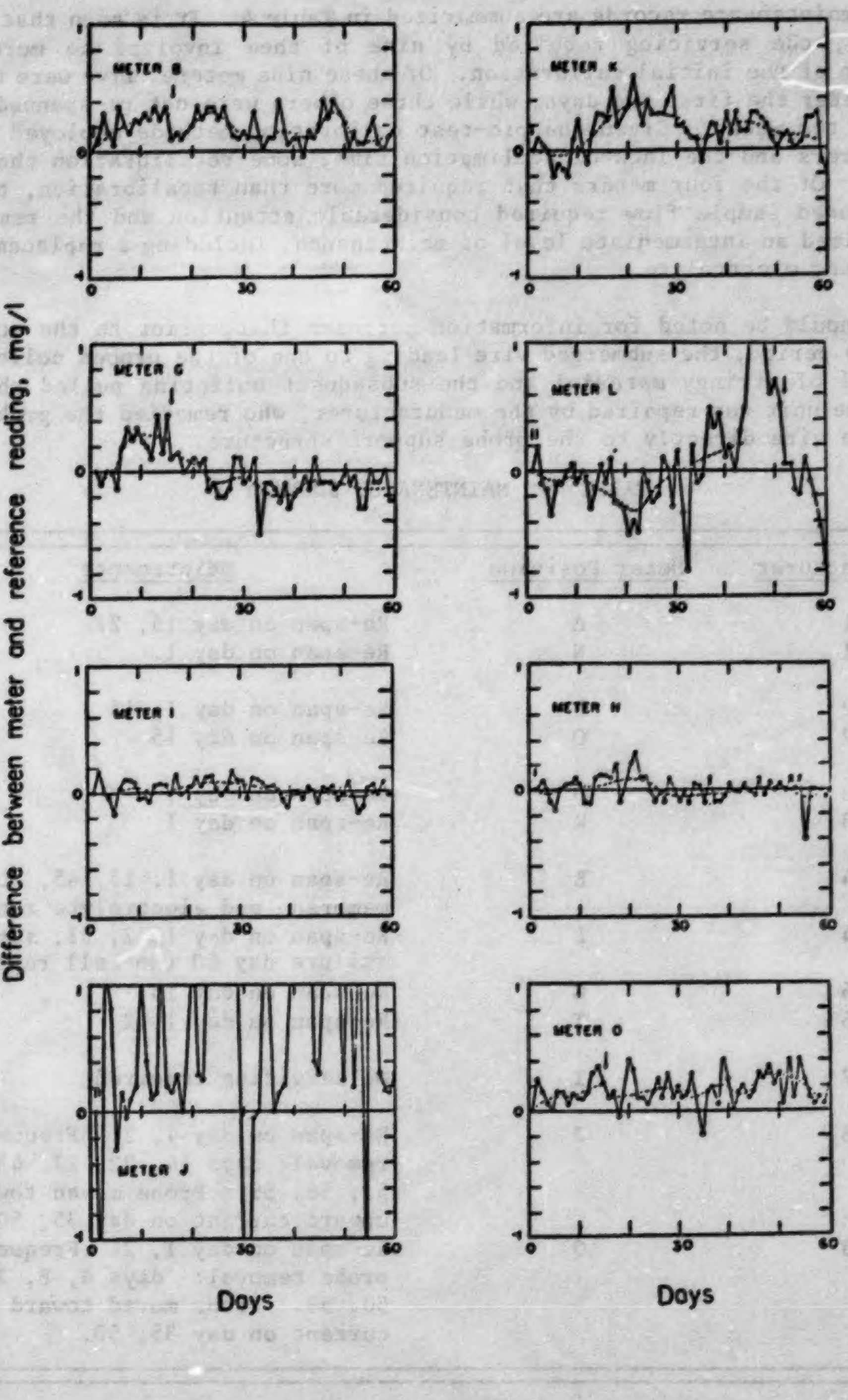

Figure 8. Examples of daily deviations.

The maintenance records are summarized in Table 4. It is seen that of the 13 meters, the servicing required by nine of them involved no more than correction of the initial calibration. Of these nine meters, five were not re-spanned after the first two days, while three others were not re-spanned after the first two weeks. Given the pre-test calibration methods employed by the manufacturers and the lack of acclimation time, some recalibration should be expected. Of the four meters that required more than recalibration, the two using induced sample flow required considerable attention and the remaining pair required an intermediate level of maintenance, including a replacement of membrane and electrolyte.

It should be noted for information purposes that, prior to the start of the 60-day period, the submerged wire leading to one of the probes collected a large ball of stringy material and the subsequent buffeting pulled the wire loose. The unit was repaired by the manufacturer, who remedied the problem by taping the wire directly to the probe support structure.

TABLE 4. MAINTENANCE SUMMARY

| Manufacturer | Meter Position | Maintenance |
|---|---|---|
| 1 | A | Re-span on day 15, 27 |
| 1 | N | Re-span on day 1. |
| 2 | b | Re-span on day 1, 15 |
| 2 | O | Re-span on day 15 |
| 3 | D | Re-span on day 1 |
| 3 | K | Re-span on day 1 |
| 4 | E | Re-span on day 1, 15, 45, 56; membrane and electrolyte replaced |
| 4 | L | Re-span on day 1, 2, 51; suspected failure day 60 (no cell replacement) |
| 6 | G | Re-span on day 15 |
| 6 | T | Re-span on day 1, 2 |
| 7 | I | No servicing required |
| 8 | J | Re-span on day 1, 2. Frequent probe removal: days 14, 22, 27, 43, 50, 51, 56, 59. Probe moved toward upward current on day 35, 50. |
| 8 | Q | Re-span on day 1, 2. Frequent probe removal: days 4, 8, 22, 27, 50, 59. Probe moved toward upward current on day 35, 50. |

NOTE: Manufacturer 5 withdrew prior to the test.

## CONCLUSIONS AND RECOMMENDATIONS

### Calibrating the Reference Meter

The use of a portable transfer standard or reference meter that is calibrated in Winkler-tested saturated clarifier effluent appears to be an accurate and controllable method for d.o. measurements in activated sludge aeration basins. The results in Table 2 show that in this case accuracy was improved by using the Winkler method to analyze the calibration solution, and it is recommended for acceptance tests or any applications in which maximum accuracy is important. For routine periodic performance monitoring, the simpler and faster air calibration is preferred provided preliminary investigations indicate that the difference between an air calibrated probe and Winkler results on clarifier effluent is not large. Caution must be taken to assure that ingredients in the plant inflow will not intermittently upset this difference. Air calibrations should be made in moisture saturated air. Our experience here also suggests that the probe should be rinsed and the calibration checked after use.

Another option involves calibrating the reference meter in distilled water that has been aerated to apparent saturation and using the standard saturation values for dissolved oxygen content. This method presumably introduced errors similar to those for air calibration (Table 2) and will require the same preliminary test when applied to effluent. On the other hand, the auxiliary tests also indicated (Table 3) that aeration with a magnetic stirrer for 20 to 30 minutes, as is frequently suggested in instructions, does not necessarily yield the tabular saturation value. This effect should be investigated if the user wishes to calibrate in this way. It is further noted in regard to both air and water calibrations that slightly different (up to about 1 percent) tabular saturation values can be found in the literature.

### Using the Reference Meter

As discussed under reference measurement development, in order to avoid momentary entrapment of air bubbles, it is recommended that the portable probe be slowly oscillated during the readings even if the sample velocities are otherwise adequate.

After the reference probe is first immersed in the sample, it must be allowed enough time (determined by monitoring the meter readings) to attain a steady state. The subsequent reference readings should be continued long enough to obtain a representative average.

The reference meter calibration should be checked after the test measurements are completed. This precaution is recommended not only to check for drift but also to guard against the results of inadvertent jostling of the meter or accidental movement of the control knobs.

### Calibrating the On-Line Meters

As indicated by the patterns in Figure 8, it is usually not meaningful to make adjustments based on the first day or two of readings, unless the deviations are substantially larger than the reference-reading error determined from estimates equivalent to those in Table 1. Rather, the deviations should be observed for at least several days while the meter equilibrates so that averages or trends can be discerned before recalibration. This applies not only to the initial test period but also throughout the test.

If the dissolved oxygen in the aeration tank varies during the day in a quasi-periodic manner, the pattern should be noted and the specific times of the daily reference readings should be varied so as to cover as large a d.o. range as possible for the performance evaluation.

Recommendations Related to Probes and Probe Mounting

The aeration basin of the activated sludge process will contain stringy fiberous material (Figure 9) as well as biological solids mixing at high velocity. The process and most equipment related to it are outside, unprotected from the weather. Maintenance and operational personnel must often work under unpleasant performance limiting conditions. At large treatment facilities repair shops and laboratories may be several hundred meters from operating equipment.

The following recommendations are designed to reduce failures caused by the above conditions, and improve the efficiency of maintenance personnel when repairs are required:

- The probe, and the submerged support structure, should be shaped to present a smooth, straight cylindrical outline to flow. That is, sharp right angle turns, abrupt enlargements, and other geometries conducive to the collection of material should be avoided. It is preferable that all wiring be confined inside the mounting elements.

- The sensor should be exposed directly to the aeration basin. That is, no sample-flow induction or agitation should be used to conduct liquid sample past the membrane. This does not apply where the sample of interest is in a quiescent state.

- Mechanical devices used to hold the probe assembly in place should be the quick release variety. That is, hand tools such as wrenches or screwdrivers should not be required to remove the probe assembly from the process. Further, fastening devices used to hold the probe assembly in place should be retained mechanically to prevent them from being dropped into the basin during maintenance.

- The design should facilitate easy, fast replacement of the membrane.

- Directly exposed sensors should be oriented so that the strongest currents do not impact perpendicularly to the membrane.

The above listed recommendations are primarily developed for shallow (1-2 meters deep) installations. Where greater submergence is required the design engineer may not be able to follow each point but should at least consider them for the specific application.

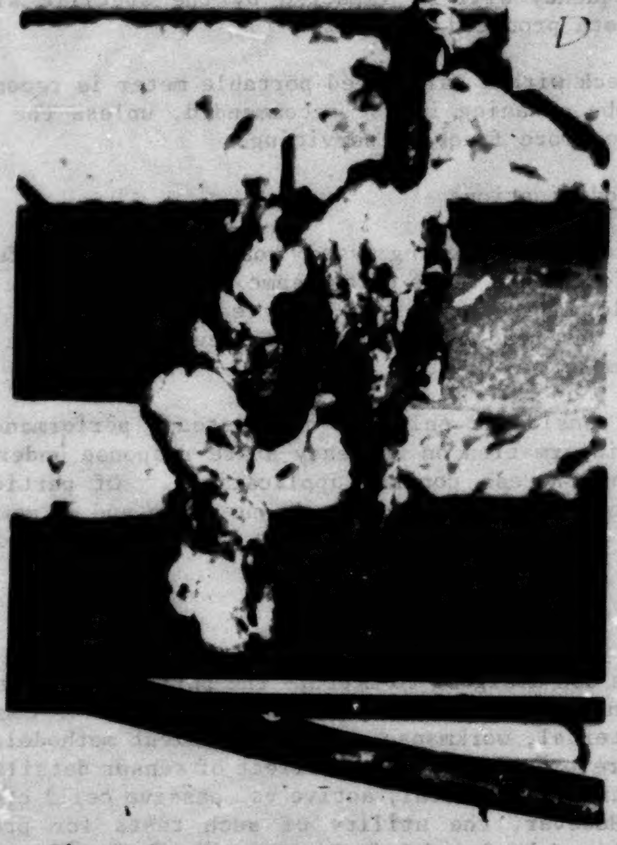

Figure 9. Example of fouled probe.

## Operation and Maintenance Recommendations

The directly exposed sensors in this test were all oriented either vertically downward or horizontally (or nearly so). Mounting these units in a flow with a general downward component proved to be a generally effective procedure, since only one case of apparent membrane rupture was experienced.

It is not possible to recommend membrane/electrolyte change intervals for preventive maintenance, because there were not enough cell failures within the 60-day period to warrant such conclusions (see Table 4). In any case, maintenance frequency will be impacted by the specific site and should be designed from data produced on-site.

A daily check with a calibrated portable meter is recommended. Membrane and general probe cleaning is not recommended, unless the daily comparisons reveal a need for more frequent servicing.

## Test Protocol Modifications

A data recording or logging system would have been useful in this test and is recommended for future tests of the same type. There were several instances in the manual logging of the four-hour readings where it was clear that an incorrect scale was read by the operator, even though the proper scale was marked on each meter.

This test considered only the quasi-steady performance of the meters. Development of information on unsteady-state response under field conditions is necessary for process control applications. Of particular interest is information on effects of cell and membrane aging and of membrane cleaning on response time. It is recommended that appropriate field test methods be developed.

## General Comment

Field tests are designed to measure the performance of a device in a given environment. Analysis of basic design criteria such as component selection, construction material, workmanship, or measurement methodology is not part of such tests. Therefore, no systematic effect of sensor details (i.e., electrode materials, electrolyte chemical, active vs. passive cell) can be inferred from these tests. However, the utility of such tests for prequalification of bidders is exhibited by O and L in Figure 8; both are passive cells using the same materials, yet there is a substantial difference in performance.

An organization wishing to conduct field tests of on-line dissolved oxygen analyzers can utilize the protocol given in the appendix provided cognizance is taken of the modifications in technical details described above. In addition, administrative details must be changed to recognize the specific organizations and locality involved in the test.

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5. "Standard Methods for the Examination of Water and Wastewater," 13th ed., APHA, AWWA, WPCF (1970).

## DISCLAIMER

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

# APPENDIX

# A FIELD EVALUATION OF ON-LINE DISSOLVED OXYGEN METERS FOR SEWAGE TREATMENT PLANTS

## 1. General

1.1 This program is proposed for the evaluation of on-line dissolved oxygen monitors under controlled conditions in a field (aeration basin) environment. Its purpose is to:

- Provide information for the development of specifications for instrument procurement at wastewater treatment plants and for acceptance testing;
- Provide information for development of accurate testing methods at wastewater treatment plants; and
- Provide information on routine maintenance requirements under a specific set of operating conditions.

1.2 This document briefly describes the proposed evaluation and lists the participants and their respective responsibilities.

### 1.3 Participants

1.3.1 The National Bureau of Standards (NBS) will provide project supervision, test procedures, reference methods or instruments, and analysis of results.

1.3.2 NBS is conducting this evaluation as part of a larger-scope interagency agreement with the Environmental Protection Agency's Municipal Environmental Research Laboratory (EPA-MERL), dealing with performance specifications and test methods for selected flow and water-quality measuring instruments in wastewater treatment plants. Program manager for NBS is Dr. George Mattingly, Chief, Fluid Engineering Division, NBS; program monitor for EPA-MERL is Mr. Walter Schuk. Immediate supervision for the dissolved-oxygen meter evaluation project will be provided by Dr. G. Kulin, NBS.

1.3.3 The Washington Suburban Sanitary Commission (WSSC) will furnish the aeration-basin location and specific support services described in this document. The WSSC contact for this program is Dr. Stanley Dea, Director, Bureau of Design.

1.3.4 Manufacturers of dissolved-oxygen meters, solicited in accordance with paragraph 2, will furnish the set up instruments for this evaluation as described in paragraph 4.

## 2. Test Instruments

2.1 **Solicitation**. Known manufacturers of dissolved-oxygen meters for the wastewater trade have been contacted by EPA-MERL representatives to ascertain their interest in participating in this evaluation. Also, the program has been announced in the Commerce Business Daily for the benefit of manufacturers who may not be known to EPA-MERL and NBS.

2.2 **Exceptions**. The agencies conducting this evaluation are committed to fairness to all manufacturers. However, in the interest of orderly investigation, the project supervisor reserves the right to reject offered instruments which do not have a demonstrable potential for activated-sludge application or which, for technical reasons, cannot be accommodated in this evaluation.

2.3 **Meters**. Manufacturers are asked to furnish two identical instrument systems, including panel-meter (or equivalent) readouts and instruction manuals. Only one pair of meters will be accepted from each source unless the manufacturer makes a second instrument of a generically different type.

## 3. General Test Conditions

3.1 **Location**. The evaluations will be conducted in an aeration basin at the WSSC Seneca Interim Wastewater Reclamation Plant on Riffle Ford Road, near Gaithersburg, Maryland. This plant uses the extended aeration process and the basin typically has 4,000 to 5,000 mg/l solids and 2 to 4 mg/l dissolved oxygen. The project supervisor will assign each instrument a specific test location within the basin. In doing this, the project supervisor will attempt to select locations such that all probes will be subjected to conditions that are sufficiently similar for the purposes of this test. Nevertheless, the probability of reasonable differences in small-scale environment is recognized and will be allowed for during the tests by appropriate monitoring and probe relocation if necessary.

3.2 **Duration**. The length of the evaluation period will be 60 days. All of the submitted instruments will be evaluated during the same 60-day period. If unforeseeable circumstances should cause an interruption of the program during this period, the conditions for (and the length of) the resumed program will be determined by the project supervisor.

## 4. Instrument installation

4.1 **Location**. The sensing probes will be mounted at the location described in paragraph 3.1.

4.2 **On-site provisions**. A standard handrail is available for mounting the equipment. The attached drawing gives pertinent dimensions and related information which enables the manufacturers to arrange for attaching their probes and analyzers to the supports. WSSC will also furnish two 110-volt outlets within 6 feet of each analyzer station.

4.3 **Installation assistance**. In general, manufacturers will be expected to provide the materials, fittings and manpower necessary to mount their equipment on the railing in accordance with paragraph 4.2 Any additional material or manpower assistance is expected to be relatively minor and should be arranged in advance with WSSC. To the extent that unusual requirements for modifications or assistance can be anticipated by the vendor well in advance, such things should be made known to the project supervisor at the time of the agreement to participate.

4.4 Installation period. A period of two weeks will be set aside during which manufacturers may set up their instruments and do all of the necessary preliminary work on them. Although the manufacturers may come in at any time during this two-week period, they will be requested to notify WSSC (Russel Sharp, Seneca Plant Superintendent, telephone 301-428-3117) of their specific installation time in advance. The two-week period will not begin earlier than June 15.

## 5. The Evaluation Tests

5.1 Start. The 60-day evaluation period will begin no later than 5 days after the end of the two-week installation period.

5.2 General test conditions. See paragraph 3.

5.3 Final adjustments. At the start of the 60-day test period, the manufacturers will be allowed to make final adjustments on their instruments if they so desire. Further adjustments will not be permitted during the test period except as noted in paragraphs 5.4 and 5.7

5.4 Maintenance. During the test period, routine maintenance will be provided by WSSC operators. The level of maintenance tasks which can be considered "routine" for each instrument will be agreed upon in discussions among NBS, EPA, WSSC and the individual manufacturers during the two-week setup period. This period may also be used by the manufacturers for communicating information and instructions on their instruments to WSSC operators and NBS personnel. Maintenance by operators will be performed only during the day shift, and then only as recommended and witnessed by the project supervisor or his representative. All maintenance that is done will be logged in detial. Maintenance requiring the services of instrumentation specialists will not be provided during the test period.

5.5 Readings.

5.5.1 Manufacturers meters. Readings with a reference meter (YSI Model 54) will be made by NBS personnel at least daily. Calibration method and calibration frequency for this meter and/or guidelines for the use of other reference methods will have been determined as described in paragraph 5.6.

5.6 Reference measurements.

5.6.1 NBS will make reference measurements with a portable dissolved-oxygen meter as noted in 5.5.2. Methods of calibrating this meter will be investigated but at this time it is anticipated that laboratory Winkler tests on clarifier effluent will be used. Prior to the start of the 60-day test period NBS personnel will also evaluate the reference meter(s) to obtain information on error band, necessary calibration frequency, methods for assuring adequate agitation, etc.

5.6.2 NBS personnel will monitor the use and calibration of the reference meter during the field evaluation and will be responsible for its maintenance.

5.7 Instrument failure.

5.7.1 For the purpose of this evaluation, "failure" or "malfunction" will include not only the obvious cases of continuous zero or continuous full-scale readings but also any continuous readings which are so constant as to constitute obvious cases of complete lack of response to change.

5.7.2 If an instrument should fail or malfunction during the test as described in paragraph 5.7.1, WSSC will take only those steps which are specifically agreed upon prior to the start of the testing and which can be accomplished by operators, and/or those steps that are normally taken by operators in these cases, such as examination for probe fouling, etc. Repairs by operators will be made only during the day shift, and then only as recommended and witnessed by the project supervisor or his representative. All repairs will be logged in detail. Repairs requiring the services of instrumentation specialists will not be made during the test period.

5.7.3 In the event of instrument failures or malfunctions which cannot be remedied by operators, the project supervisor will decide on a case-by-case basis whether a particular instrument should be withdrawn from further participation in the tests or whether the manufacturer should be asked to repair/replace it. These decisions will attempt to take into account both the need for useful test data and the commitment to fairness to all manufacturers. In cases where replacement is permitted, only an identical model of the instrument will be acceptable. In any event, manufacturers will be asked to determine the cause of failure and notify the project supervisor for information purposes.

5.7.4 If the failure of an instrument is due to an obvious accident or incident beyond the control of the manufacturer and not normally encountered in activated sludge basin, repair or replacement of the instrument and continuation of its evaluation will be permitted, but the final determination will be made by the project supervisor.

# THE CALIBRATION AND EVALUATION OF DISSOLVED OXYGEN SENSORS IN A PILOT SCALE ACTIVATED SLUDGE PLANT

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Burlington, Ontario L7R 4AD

## ABSTRACT

The performance of three types of dissolved oxygen (DO) sensors was evaluated in a pilot scale activated sludge plant. As the instruments were interfaced to a minicomputer, a curve fitting technique was developed to compensate for instrument non-linearity. Weighted regression was performed to update initial calibrations.

Fouling due to the microbial growth on the surface of the probe was the principal cause of decline in sensor performance with time. The rate of fouling depended on the type of sensor used, the mode of operation, the frequency of maintenance and the velocity of fluid past the probe. A number of alternatives were evaluated for reducing or compensating for the effects of fouling and other causes of inaccuracy.

To determine the suitability of sensors for tracking DO concentration during non-steady state tests for determining overall mass transfer coefficients ($K_La$), the probes were subjected to repeated step changes in DO concentration. First-order time constants ($\tau$) for the sensors were estimated based on the results of each step.

## INTRODUCTION

For aerobic wastewater treatment systems it is desirable to obtain, on a continuous basis, an accurate indication of dissolved oxygen concentration in the aeration compartment. Electrochemical dissolved oxygen monitors are readily available and used routinely to aid in the maintenance of a minimum concentration of dissolved oxygen necessary to insure the viability of the microbial population, to allow blowers and aerators to be controlled to provide adequate but not excessive aeration and, finally, to permit aeration equipment to be evaluated under field and laboratory conditions. Experience gained in the continuous on-line monitoring of dissolved oxygen concentration with different makes of sensors resulted in the development and evaluation of methods for improving sensor performance. Factors influencing the suitability and performance of DO sensors included installation, calibration, maintenance and the dynamics of the sensor in responding to sudden changes in DO level.

## RESEARCH EQUIPMENT

### Pilot Plant Facility

The research was carried out using an extended aeration activated sludge package plant modified and operated in the conventional mode. Flow to the plant, controlled at a rate of 200 $m^3/d$, consisted of degritted municipal wastewater. The aeration basin was baffled into three sections, each with a volume of 22.2 $m^3$. Air was supplied through coarse bubble diffusers by a 5.6 kW positive displacement blower. On-line process instruments interfaced to a real-time minicomputer provided monitoring and control for the plant.

### Dissolved Oxygen Sensors

Dissolved oxygen sensors from three different manufacturers; Yellow Springs Instrument Company (YSI), pHOX Systems Ltd., and Zullig, were used in this research. Each consisted of a probe and an analyzer/transmitter which displayed the indicated dissolved oxygen reading in milligrams dissolved oxygen per liter.

Commercial dissolved oxygen probes are electrochemical cells which generally operate on one of two principles. Galvanic cells consist of a noble metal cathode, a base metal anode and an electrolyte. Chemical reactions involving oxygen ojcur at the cathode and are spontaneous, causing a measurable current flow in proportion to the oxygen concentration present. Polarographic cells operate in a similar manner but use two noble metal metal electrodes and require a constant external voltage to polarize the electrodes before the necessary chemical reactions occur. In some cases, a gas-permeable plastic membrane isolates the sensor elements and electrolyte from foreign material and impurities in the sample. Usually the electrolyte is potassium hydroxide, potassium chloride or potassium bicarbonate in liquid or gel form. For some makes of DO sensors the electrodes are in direct contact with the sample stream which serves as the electrolyte. Partial specifications for the three makes used in this research are listed in Table 1 and each probe is shown in Fig. 1.

Table 1. List of instrument specifications

| Make and Model | Number Evaluated | Range* (mg/l) | Operating Principle | Membrane Type | Electrodes | Electrolyte | Temperature Compensation |
|---|---|---|---|---|---|---|---|
| YSI<br>Model 54A Transmitter<br>Model 5734 Probe | 1 | 0-20 | Polarographic | Teflon | Gold<br>Silver | Liquid<br>KCl | Yes |
| pHOX<br>Model 66 Transmitter<br>Model 65-88 Transmitter<br>Model 670 Probe | 3 | 0-20 | Galvanic | Polythene | Silver<br>Porous Lead | Gel<br>$KHCO_3$ | Yes |
| Züllig<br>Model P-76-2-1 | 1 | 0-20 | Galvanic | – | Amalgam<br>Zinc | – | Yes |

* Instruments were used at the 0-10 mg/l scale setting.

The YSI dissolved oxygen sensor is a portable unit commonly used for analytical procedures and discrete sampling at wastewater treatment plants. A thin teflon membrane is stretched over the cell electrodes located at the end of the probe. The membrane and electrolyte are easily replaced by the user. The unit is ideal for comparing calibration of other sensors and one was used as a reference exclusively for this purpose. A second unit was used for continuous on-line monitoring in a dip mode, however, additional electronic signal isolation equipment was required to produce acceptable results.

The pHOX probe is of robust construction and designed for continuous on-line monitoring in either a dip or flow-through mode. A perforated, cylindrical silver cathode of large surface area fully supports a tubular polythene membrane. The membrane is replaceable but the procedure requires a great deal of care. The cell is a plug-in unit which, when exhausted, is rejuvenated by the manufacturer on an exchange basis.

The Zullig sensor uses a dip probe designed to be self cleaning. A motor drives an electrode-cleaning grindstone and activates a beaker surrounding the electrodes to alternately draw and expel a fluid sample on a continual basis. Since the sample serves as the cell electrolyte, this action ensures that the sample is constantly renewed to maintain conductivity. The unit is rugged and designed for continuous on-line use. The grindstone is the only user-replaceable part and is easily changed.

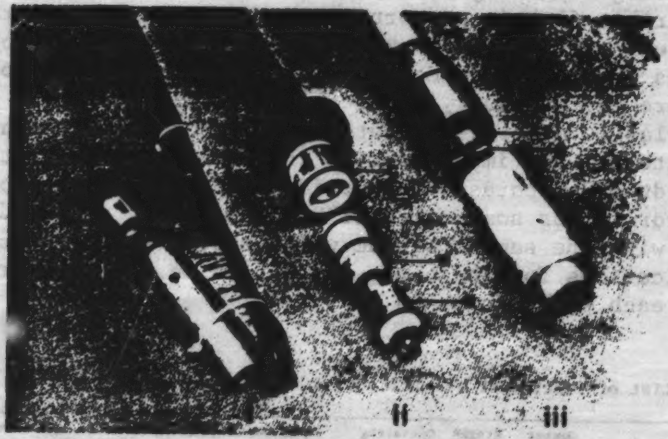

Figure 1. Dissolved oxygen probes: i) YSI; a) protective shield, b) electrode head. ii) pHOX; c) probe head, d) porous lead anode, e) silver, cylindrical cathode. iii) Züllig; f) electrode assembly, g) rotating grindstone, h) cam-action draw and fill beaker assembly

## Sensor/Computer Interface

All of the DO sensors, with the exception of the reference YSI were interfaced to a real-time minicomputer. As shown schematically in Figure 2, the DO concentration was converted by the DO sensor to an analog signal

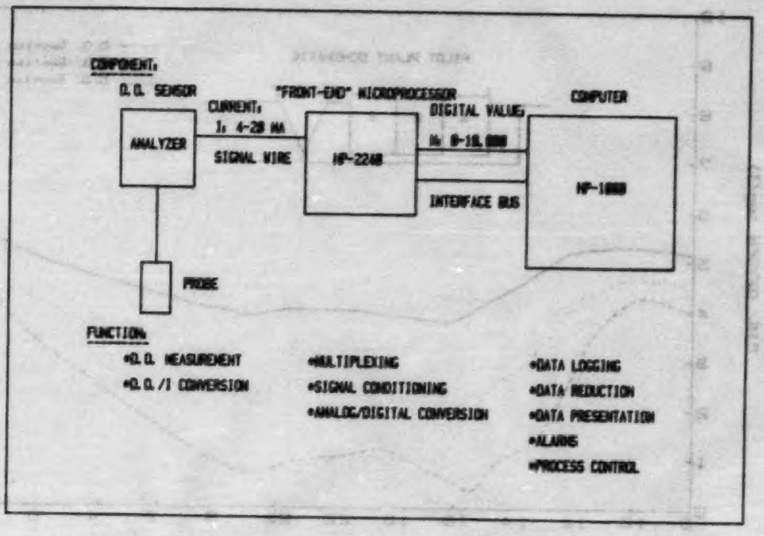
COMPONENT:
D.O. SENSOR
"FRONT-END" MICROPROCESSOR
COMPUTER
CURRENT:
I: 4-20 MA
SIGNAL WIRE
ANALYZER
HP-2240
DIGITAL VALUE:
N: 0-10,000
INTERFACE BUS
HP-1000
PROBE
FUNCTION:
•D.O. MEASUREMENT
•D.O./I CONVERSION
•MULTIPLEXING
•SIGNAL CONDITIONING
•ANALOG/DIGITAL CONVERSION
•DATA LOGGING
•DATA REDUCTION
•DATA PRESENTATION
•ALARMS
•PROCESS CONTROL

**Figure 2. Components of computer monitoring system**

which was sampled, conditioned and digitized by the front-end microprocessor. The digitized sensor signal was converted by computer software to appropriate engineering units. As the lowest value for the signal output from the computer was 20% of the signal range, the computer could distinguish between an actual measured value of zero DO concentration - a "live zero" - and a zero caused by the instrument being inadvertently switched off or by a damaged signal wire. An alarm condition (the DO concentration below a critical minimum) if detected by the computer would sound a buzzer to alert the plant operator and write a message describing the nature of the alarm to a terminal.

Data collected from DO sensors were summarized and displayed using computer peripherals including terminals (CRTs), a line printer and a plotter. For normal data logging, the sensor signal was sampled at a five-second interval with the data stored in disc files as 15-minute averages. Figure 3 shows a plot of DO for three sensors, each placed in one of the compartments of the aeration basin. For a constant airflow rate, the plot illustrates the effect of diurnal loading on DO concentration as well as the variation in DO for three complete mix reactors in series.

## CALIBRATION PROCEDURES

### Standard Calibration Procedure

Accurate calibration of DO sensors is essential for reliable process monitoring and control. Although specific calibration procedures vary with make and model of instrument, DO sensors are generally calibrated by adjusting the instrument zero and span.

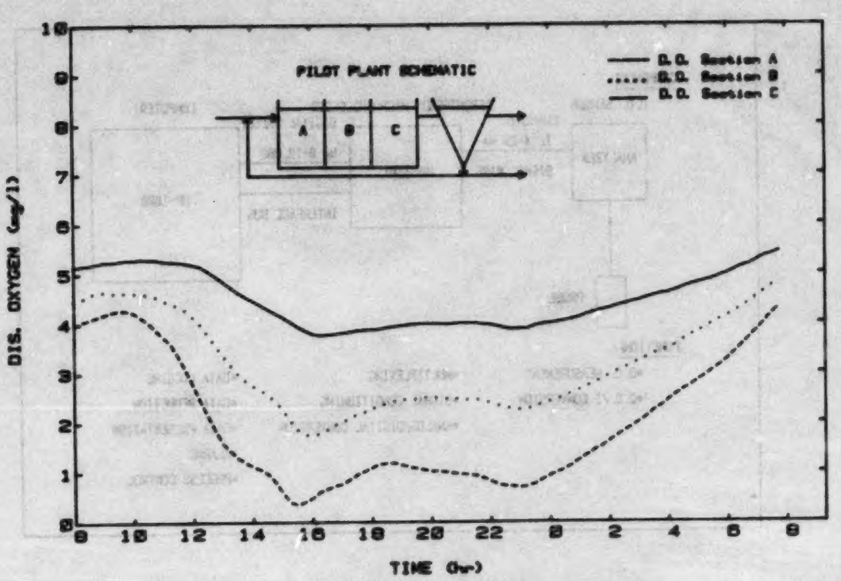

Figure 3. Changes in DO concentration due to diurnal loadings

Adjustment of the instrument zero, if provided, can be accomplished by manipulating the zero potentiometer with the selector switch at the zero position. With the selector switch in this position, the instrument is not influenced by the signal received from the probe. Alternatively, the instrument zero is adjusted using a DO-free solution obtained by adding excess sodium sulfite and a trace of cobalt chloride to water (Standard Methods, 1975). The instrument span is adjusted following the immersion of the probe into a bucket of water which has been aerated to saturation. The dissolved oxygen content of the sample is determined according to saturation tables or accepted analytical methods. For sensors of the membrane type, some manufacturers suggest air calibration whereby the electrode is allowed to reach equilibrium in air. The span is adjusted to match the value obtained from saturation tables.

A number of calibration details influence the resulting sensor accuracy. The calibration samples must be stirred to eliminate stagnant layers of water which interfere with the transfer of oxygen into the electrochemical cell. For sensors with electrodes in direct contact with the water and, therefore, without membranes or electrolyte, the calibration samples must be sufficiently conductive for the probe to register DO. For example, for the Züllig sensor, a minimum of 300 μsiemen was recommended for calibration. Interaction between zero and span potentiometers requires repeated checks at the high and low DO levels to ensure accuracy.

## Regression Techniques as Applied to Sensor Calibration

The calibration procedure described in the preceding section fails to detect or compensate for non-linearity in DO sensors. Sensor non-linearity, sometimes expressed as a percentage, is the amount of deviation of the relationship between DO concentration and measurement signal from a straight line defined by the measurement signal at zero DO and at maximum DO (Liptak, 1969). As DO sensors were interfaced to a minicomputer, a more sophisticated calibration procedure could be implemented using regression analysis to compensate for detected non-linearity.

Data pairs consisting of a reference YSI DO reading, $y_i$, and corresponding digital value, $x_i$, of the test sensor were collected over the range of the test sensor. To accomplish this, the probes were immersed in a sample of mixed liquor which had been aerated to saturation. The data pairs, $x_i$ and $y_i$, were collected at approximately 0.5 mg/l intervals as the DO level decreased due to microbial respiration.

The data pairs $(x_i, y_i)$ were plotted to determine the degree of non-linearity. When the digital values varied linearly with the reference YSI readings, a calibration equation of the form $\hat{y} = a + bx$, based on simple linear regression, was used for the software conversion of incoming signals. For a relationship between the digital values and DO, which was non-linear, a curve or combination of lines and/or curves was fitted to the data.

Experience over several months indicated that, for non-linear cases, it was necessary to fit, at most, two lines or a line and a cuuve to the observed values. For those portions of the plots which were smooth curves, a number of functions were used to fit the data. The most convenient functions were those which could be transformed to a linear form of the type $\hat{y}' = a + bx'$ where $y'$ and $x'$ are defined in Table 2. Functions were fitted to the curve by carrying out the appropriate transformation on the data and performing linear regression on the transformed data. A summary of a number of functions and the appropriate transformations is given in Table 2.

The results of the regression technique for two sensors which exhibited non-linearity are given in Fig. 4 and Fig. 5. The calibration relationship plotted in Fig. 4 was represented by two straight lines. For Fig. 5 the calibration relationship was represented by a line and a geometric curve.

The regression technique as applied to sensors has a number of advantages when compared with less systematic approaches to calibration. They include:

a) detection of and compensation for sensor non-linearity;
b) significant reduction in the amount of adjustment required should the zero and span be interactive or the internal potentiometers require adjustment to restore sensor linearity; and
c) ability to quantify and document changes in probe characteristics with time.

**Table 2. Curvilinear functions of one variable and transformations to linear form (after Himmelblau, 1968)**

| Equation | Transformation ($y' = a + bx'$) | |
|---|---|---|
| | $y'$ | $x'$ |
| $1/y = a + bx$ | $1/y$ | $x$ |
| $y = a + b(1/x)$ | $y$ | $1/x$ |
| $x/y = a + bx$ | $x/y$ | $x$ |
| $y = ax^b$ | $\log y$ | $\log x$ |
| $y = ab^x$ | $\log y$ | $x$ |

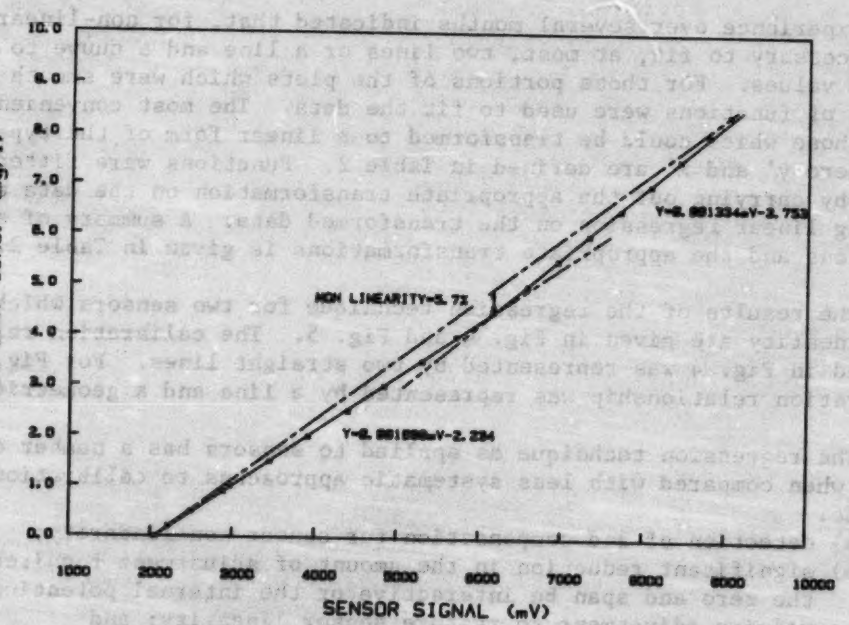

**Figure 4. Sensor non-linearity - two straight lines**

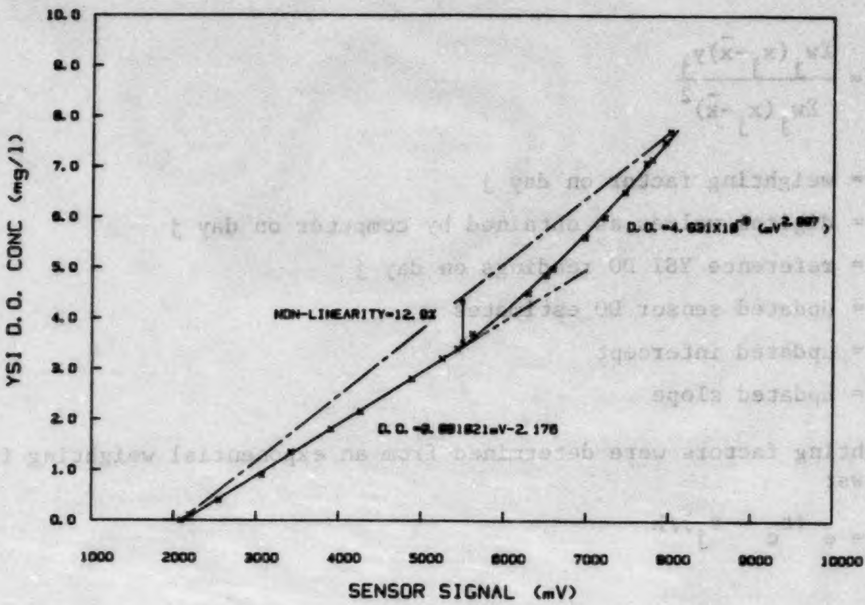

**Figure 5. Sensor non-linearity straight line and geometric curve**

## Updating the Calibration Equation

To counteract gradual changes in instrument performance which with time invalidated the initial calibrations, paired reference readings were added to the initial calibration data equations. As paired values obtained more recently were more credible than "older" calibration data, the updating procedure weighted the data accordingly. Weighted regression was used to obtain the updated calibration equations as follows (Himmelblau, 1968):
Updated equation:

$$\hat{y} = a_u + b_u x$$

where:

$$a_u = b_0 - b_u \bar{x}$$

$$\bar{x} = \frac{\Sigma w_j x_j}{\Sigma w_j}$$

$$b_0 = \bar{y} = \frac{\Sigma w_j y_j}{\Sigma w_j}$$

$$b_u = \frac{\Sigma w_j (x_j - \bar{x}) y_j}{\Sigma w_j (x_j - \bar{x})^2}$$

$w_j$ = weighting factor on day j

$x_j$ = digital values as obtained by computer on day j

$y_j$ = reference YSI DO readings on day j

$\hat{y}$ = updated sensor DO estimates

$a_u$ = updated intercept

$b_u$ = updated slope

The weighting factors were determined from an exponential weighting function as follows:

$$w_j = e^{-(t_c - t_j)/k}$$

where:

$w_j$ = weighting factor at $t_j$; $0 \leq w_j \leq 1.00$

$t_c$ = current day

$t_j$ = day on which the data pair $x_j$, $y_j$ were obtained

k = constant

The weighting function dictated that data obtained on the current day counted 100% in the regression analysis performed on that day; data obtained k days prior to the current day was counted 37%; data obtained 2 k days prior counted 14% and so on. The constant k can therefore be thought of as a "pseudo-time constant" which determines the influence of past data on the updated calibration equation. The results of the technique are illustrated in Fig. 6 and Fig. 7. Figure 6 shows the initial calibration relationship of February 3 and seven subsequent values obtained to update the calibration.

Using an arbitrarily selected k value of 7.0 days and the data of Figure 6, the lines representing the plots of the updated calibration equations are shown in Figure 7.

To compensate for membrane fouling, the calibration line for February 24 was shifted, as a result of the regression analysis, to a position above the initial calibration line. The probe was cleaned on February 25 and the subsequent calibration lines (February 26 and March 2) were returned to positions close to the initial calibration line. The weighted regression technique updated the initial calibration with more current information, essentially "tuning" the sensor.

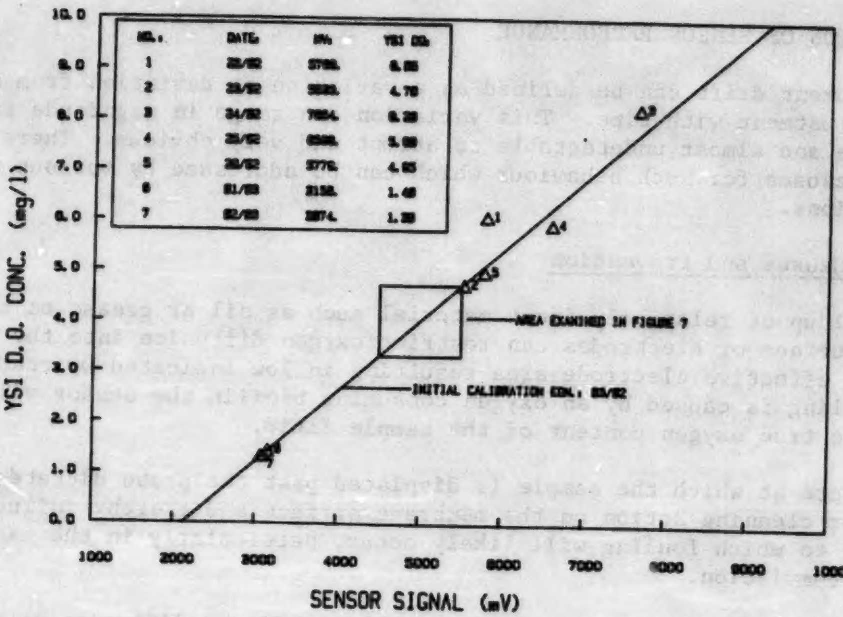

Figure 6. Weighted regression initial calibration relationship with subsequent data

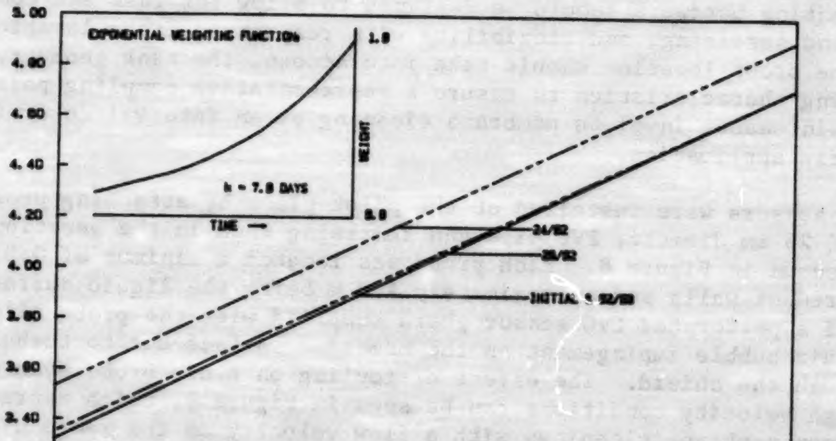

Figure 7. Weighted regression: updated calibration relationships

## DETERIORATION OF SENSOR PERFORMANCE

Instrument drift can be defined as a variation or deviation from an initial adjustment with time. This variation can range in magnitude from very subtle and almost undetectable to abrupt and very obvious. There are a number of causes for such behaviour which can be addressed by various preventative actions.

### Fouling: Causes and Prevention

A buildup of relatively inert material such as oil or grease on the membrane surface or electrodes can restrict oxygen diffusion into the cell or reduce the effective electrode area resulting in low indicated DO readings. If the fouling is caused by an oxygen consuming biofilm the sensor will not measure the true oxygen content of the sample fluid.

The rate at which the sample is displaced past the probe dictates the scouring or cleaning action on the membrane surface and thereby influences the extent to which fouling will likely occur, particularly in the case of biofilm accumulation.

Aeration systems which use coarse bubble submerged diffusers or mechanical surface aerators will likely induce adequate *in situ* fluid velocities to maintain the necessary cleaning action. Under such relatively high velocity conditions, dip probes can be exposed directly to the aeration basin. This installation mode is relatively simple and requires little additional equipment. Mounting hardware should be designed to allow for fast and convenient cleaning and servicing, and flexibility with respect to probe location in the basin. The probe location should take into account the tank geometry and fluid mixing characteristics to ensure a representative sampling point. Routine maintenance involves membrane cleaning at an interval determined for the specific application.

pHOX sensors were installed at the pilot plant by attaching probes to lengths of 25 mm diameter PVC pipe and immersing them in the aeration compartments as shown in Figure 8. Each probe was located a minimum of 0.5 m from the compartment walls and approximately 1.0 m below the liquid surface. Removal of a perforated PVC sensor guard supplied with the probe eliminated possible air bubble impingement on the membrane surface due to turbulent flow through the shield. The effect of fouling on a dip probe even under fairly high velocity conditions can be seen in Figure 9, which represents 21 days between membrane cleanings with a flow velocity in the range of 0.3 to 0.6 m/sec.

Stephenson *et al*. (1981) found that dip probes used at low velocity conditions rapidly became fouled and required high maintenance to ensure good performance. Two alternatives available for low velocity conditions include the inducement of fluid flow past the probe and the use of self-cleaning probes. A pHOX probe was installed in a flow through cell (see Fig. 10) through which mixed liquor from the aeration basin was circulated by a mechanical pump. The inlet and outlet piping arrangement resulted in flow

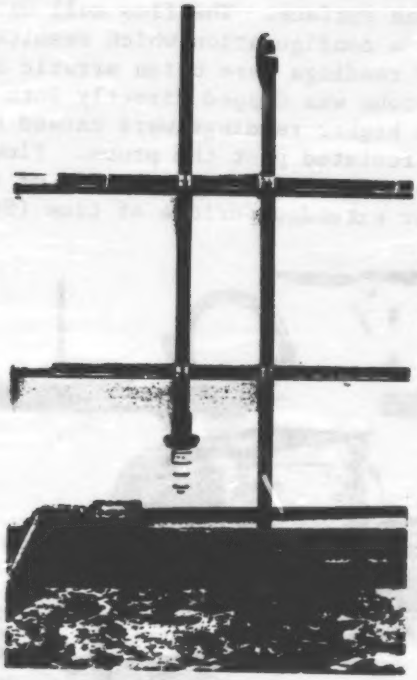

Figure 8. Dip mode showing mounting hardware

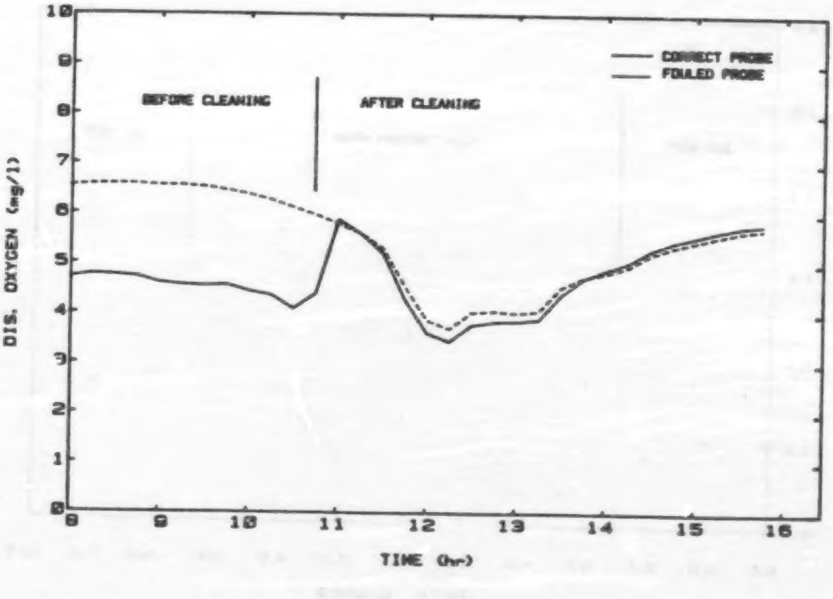

Figure 9. The effects of probe cleaning on sensor performance

tangential to the membrane surface. The flow cell was located on the discharge side of the pump, a configuration which resulted in unsatisfactory performance since sensor readings were often arratic and higher than those obtained when the same probe was dipped directly into the aeration basin as shown in Figure 11. The higher readings were caused by air entering the pump casing and being circulated past the probe. Flow cells used on other pilot plants have been placed on the suction side of the pump and have performed satisfactorily for extended periods of time (Stephenson _et_ _al_., 1981).

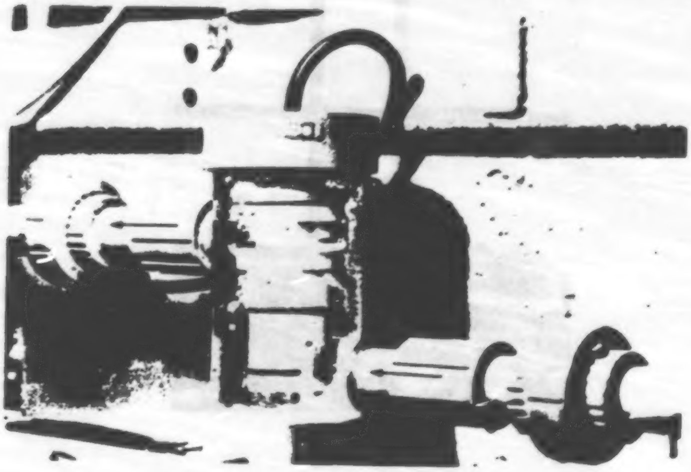

Figure 10. Flow through configuration

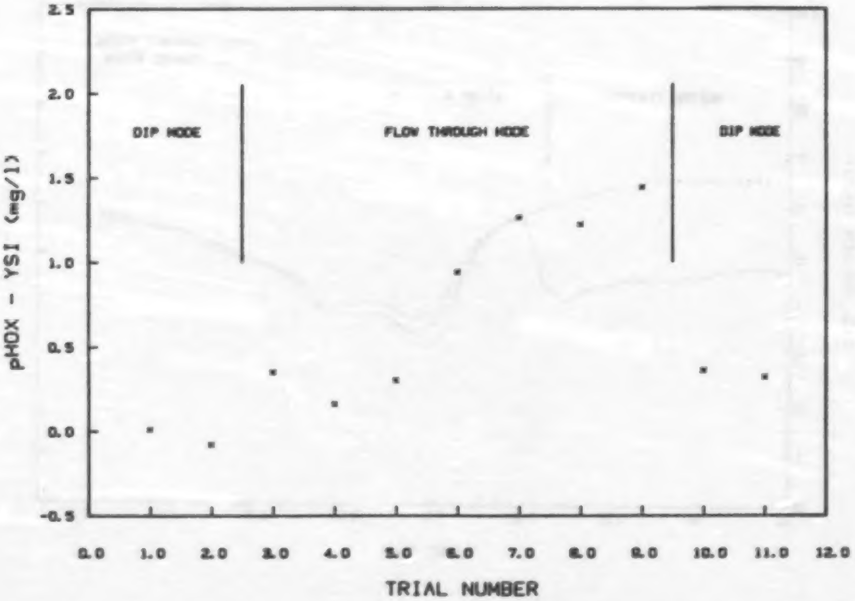

Figure 11. Difference in sensor readings according to installation mode

A comparison of the two installation modes suggest that because of the ease of inspection and cleaning and the flexibility of location, the dip mode be used where velocities provide sufficient cleaning action. Flow inducers require additional hardware and equipment (and, therefore, create other maintenance problems) but reduce membrane fouling.

The self-cleaning Züllig required periodic servicing of the grinding mechanism. The grindstone became "blinded" with material from the electrodes which resulted in less effective cleaning and arratic performance. As documented in Figure 12, this can lead to a rapid drop in indicated dissolved oxygen and should be regularly checked to ensure data integrity.

Other causes of performance loss relate to physical deterioration of equipment. Electrode decomposition and electrolyte inactivity require that the probe be returned to the manufacturer for repair and rejuvenation or replacement. Membranes age, gradually suffer loss of permeability, and must be replaced. The analyzer/transmitter, like other electronic instruments, is full of electronic components which are subject to failure. Trouble shooting, adjustment and repair can be extremely time consuming and requires skilled and experienced technical personnel.

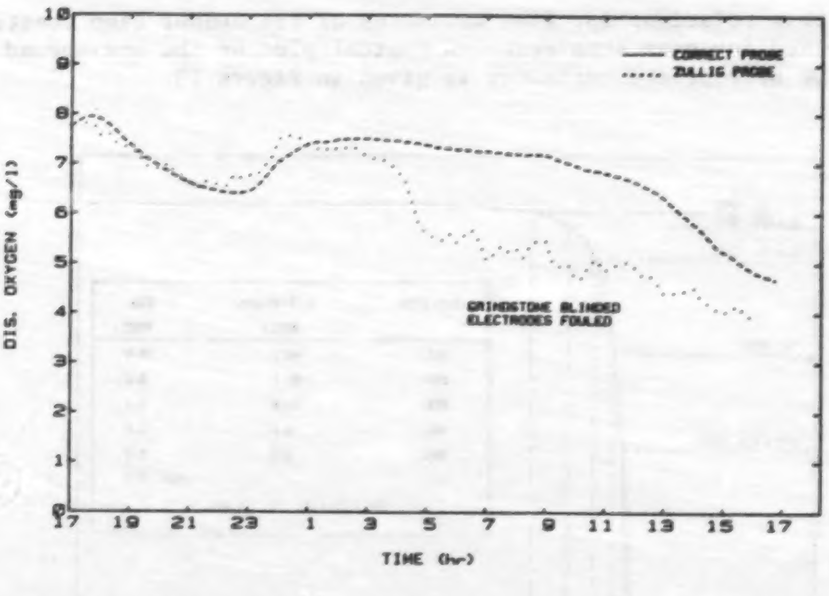

Figure 12. Plot demonstrating failure of the self cleaning mechanism

## PROBE DYNAMICS

One important application for DO sensors is the monitoring of DO concentration during non-steady state tests for determining overall mass transfer coefficients ($K_L a$). The suitability of DO sensors for tracking DO concentration which was changing with time was determined by measuring a first-order time constant ($\tau$) using step testing.

An upward step in DO was induced by transferring the DO probe under investigation from a stirred bucket containing DO-free water to a second stirred bucket containing DO-saturated water. A downward step was induced by reversing the order of transfer. The transfer was accomplished in less than one second. During the step testing, the output signal from the DO sensor was sampled by the computer at one second intervals.

For a first-order system responding to a step change, the following relationship applies between the time elapsed following introduction of the step and the percentage of the total step completed:

| Elapsed Time | Percent of Total Step |
|---|---|
| 1/2 x τ | 39.3 |
| τ | 63.2 |
| 2 x τ | 86.5 |
| 3 x τ | 95.0 |
| 4 x τ | 98.2 |

where τ = system first-order time constant.

Using this relationship, five estimates of the sensor time constant were determined for each step test. A typical plot of the one-second values with the five appropriate estimates is given in Figure 13.

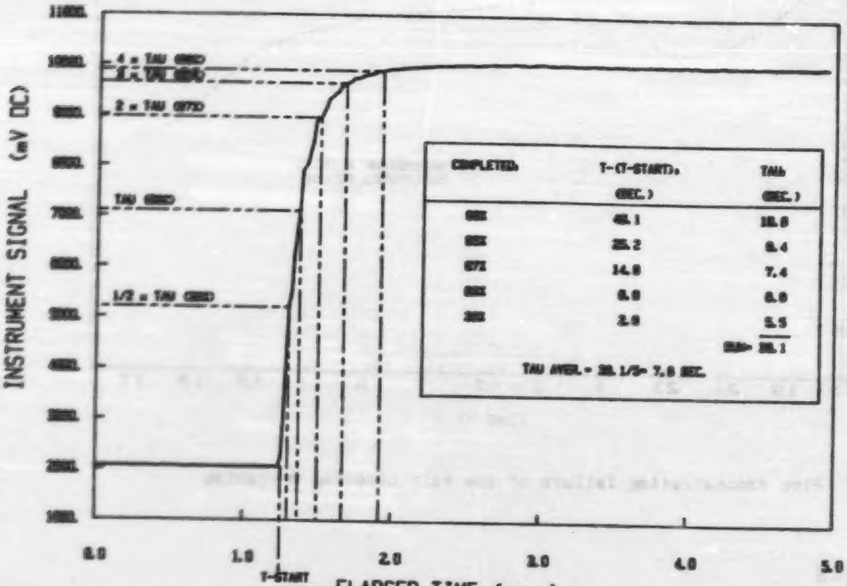

**Figure 13.** Plot of sensor response to a step change in DO and time constant estimates

The step test results are summarized in Table 3 which contains, for each of the steps, values of the first-order time constant (τ) estimated at 63% of the response, 95% of the response and as the mean of the estimates at 39%, 63%, 87%, 95% and 98% of the response. Also included in the table is an estimate for a time constant for a Beckman analyzer and Model 39553 DO sensor as measured by Shiba (1981).

Table 3. Sensor first-order time constant as measured by step testing

| | Temperature of Solution | | | Time Constants (τ) | | |
|---|---|---|---|---|---|---|
| | DO Free (°C) | DO Saturated (°C) | Step | At 63% (sec) | At 95% (sec) | Mean (sec) |
| pHOX (1,1)* | 18 | 17 | Up | 6.8 | 8.4 | 7.6 |
| | 15 | 15 | Up | 6.6 | 8.1 | 7.4 |
| | 17 | 15 | Down | 8.5 | 15.3 | 13.3 |
| pHOX (1,2)* | 16 | 13 | Up | 14.6 | 23.1 | 20.1 |
| | 13 | 13 | Down | 11.9 | 22.5 | 18.5 |
| | 16 | 13 | Down | 10.5 | 20.4 | 17.3 |
| pHOX (4,9)* | 16 | 15 | Up | 8.1 | 19.6 | 15.6 |
| | 15 | 14 | Up | 9.9 | 18.9 | 15.8 |
| | 16 | 14 | Down | 13.9 | 25.9 | 22.0 |
| | 15 | 15 | Down | 11.3 | 20.9 | 17.3 |
| Züllig | 15 | 14 | Up | 27.9 | 17.0 | 23.9 |
| | 13 | 13 | Up | 10.9 | 10.2 | 10.7 |
| | 15 | 14 | Down | 14.4 | 14.2 | 15.3 |
| | 15 | 15 | Down | 14.5 | 13.5 | 14.6 |
| YSI | 21 | 24 | Up | 8.5 | 13.9 | 11.5 |
| | 21 | 23 | Up | 7.6 | 10.4 | 10.1 |
| | 23 | 23 | Down | 13.0 | 17.3 | 15.1 |
| | 21 | 24 | Down | 12.8 | 15.6 | 14.5 |
| Beckman | | | Up | 15.0** | - | - |

* The numbers refer to the slot and channel of the sensor interfaced at the computer. They are used to distinguish between the three sensors of the same make.

** After Shiba, 1981.

For all the sensors, with the exception of the Züllig (the only probe without a membrane or electrolyte), the estimates of the first-order time constant increased as the percentage of the total response (upon which the estimate was based) increased. Therefore, for the sensors employing membranes, the dynamic response is not strictly first-order. The deviation from first-order could be caused by diffusion limitations across the membrane.

A Student's t-test was performed to determine if a difference in time constant could be detected between responses to upward and downward steps. The YSI was the only probe for which there was a significant difference at the 95% confidence level. Based on the results of the t-test, the mean first-order time constants are summarized in Table 4.

As can be seen in Table 4, the mean time constants were less than 20 sec. and were in the same range as for the YSI which would normally be used for DO determination during non-steady state $K_La$ tests. Therefore, based on the results of the step tests, the on-line probes were judged suitable for monitoring DO concentration during non-steady state $K_La$ tests.

The dynamic response of a DO sensor as measured by step testing is thought to be influenced by such factors as the size of the forcing step, test conditions, and the age and condition of the membrane and cell. Kok and Zajic (1975) determined YSI response time for a number of downsteps and found that the response time was not influenced by the magnitude of the downstep over the range investigated. Additional work is required before this conclusion can be applied to other makes. Kok and Zajic also observed a decrease in the YSI time constant from one day to the next and attributed the change to the aging and stretching of the membrane. These observations and comments by other researchers (Kulin and Schuk, 1978) indicate that membrane and cell aging may exert a considerable influence on sensor response time and that further research is justified to expand knowledge in this area.

Table 4. Summary of first-order time constants

| Sensor | Mean First-Order Time Constants (sec) |
|---|---|
| pHOX (1,1) | 9 |
| pHOX (1,2) | 19 |
| pHOX (4,9) | 18 |
| Züllig | 16 |
| YSI | 11 (up) |
|  | 15 (down) |

## SUMMARY AND CONCLUSIONS

Experience with sensors used to continuously monitor dissolved oxygen concentration in the aeration tank of an activated sludge pilot plant resulted in improved calibration procedures, evaluation of installation and maintenance alternatives and determination of first-order time constants for each of the sensors.

Standard calibrations which involved adjustment of instrument zero and span failed to detect or compensate for non-linearity. Regression analysis on pairs of values from the reference and test sensors compensated for non-linearity, reduced the number of potentiometer adjustments and provided documentation of the calibration process. Exponentially weighted regression analysis used more recently obtained reference values to update the initial calibration equation and correct for drift.

Probe fouling due to the growth of microbial films was the major maintenance problem. The severity of the problem depended on the fluid velocity in the aeration tanks. For basins with high fluid velocities, probes immersed directly into the liquid performed well and provided flexibility and simplicity of installation and inspection. With low fluid velocities, special "self-cleaning" probes, flow-through cells which use pumps to convey the fluid past the probe, or frequent inspection and cleaning of probes used in the dip mode, were required to maintain performance.

Sensor inaccuracy also resulted from membrane aging and damage, the clogging of the anode and cathode with decomposition products, electrolyte exhaustion and failure of electronic components. These problems of physical deterioration occurred less frequently and are a function of the design life of the sensor components.

First-order time constants were determined by subjecting each of the sensors to repeated upward and downward steps in DO concentration. The mean time constants for the probes tested were estimated to be less than 20 sec. in all cases. The dynamic response of a sensor without a membrane was much closer to that of a first-order system than was that of sensors with membranes. Additional research is required to determine the effect of cell and membrane aging on the dynamic response of DO sensors.

It is hoped that this work will encourage other users to communicate their experiences with installation, calibration and maintenance of sensors used to continuously monitor or control DO in the activated sludge system.

## ACKNOWLEDGEMENTS

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## DISCLAIMER

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# OPERATION AND MAINTENANCE OF CERAMIC FINE BUBBLE DIFFUSERS IN ACTIVATED SLUDGE SYSTEMS

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## ABSTRACT

Experience with the use of ceramic fine bubble diffusers for activated sludge aeration in the United Kingdom, Holland, and the U.S. is discussed. Results of a survey of 19 treatment plants using this equipment and the author's subsequent experience and observations are summarized.

The impact of plant design on the subsequent efficiency and operability of ceramic diffuser systems is outlined. Causes of failure and their symptoms are summarized. Design and operational strategies for optimal system performance are introduced.

## INTRODUCTION AND BACKGROUND

It has been recognized for some time that fine bubble aeration systems for activated sludge are inherently more energy efficient than coarse bubble systems. The use of fine bubble aeration systems for oxygenation of activated sludge processes has been practiced with varying results since around the turn of the century. Earlier systems often used ceramic plates which were mounted in place on the tank floor over a cast in place air distribution manifold. Cleaning of these systems was costly and time consuming. The steady decline of energy costs in real dollars after World War II, coupled with the development of very low maintenance coarse bubble diffuser systems resulted in the conversion of many of the earlier fine bubble systems in the U.S. and use of coarse bubble equipment in most new plants.

Historically, energy costs in the United Kindom and Europe have been considerably higher than those in the U.S. Consequently, there has been continued development and application of more efficient aeration systems there, particularly in the U.K. Slow speed mechanical turbines, typlified by the Simplex and Simcar aerators, were developed as a more efficient alternate to coarse bubble systems and were widely applied in the U.K., and later in the U.S. In the late 1950s, Activated Sludge Ltd., developed and patented a fine bubble ceramic diffuser using an air grid that was mounted above the aeration tank floor. The individual cylindrical diffusers were about 6 inches high and 4 inches in diameter and were mounted on top of the air pipes in an even pattern. The entire assembly was bolted to the tank floor. The intent of this early design was to greatly simplify the construction and maintenance of ceramic fine bubble diffuser systems while retaining whole tank floor coverage. A second generation of the ceramic dome diffuser was brought out in the late 1960s. The height of the diffuser was reduced to about 2 inches and the diameter was increased to 7 inches resulting in an improved bubble pattern and greater capacity. This design, shown in Figure 1, is the predominant type of ceramic fine bubble diffuser in use in England, the U.S., and Canada and is manufactured by Hawker Siddeley Ltd. in the U.K. and by the

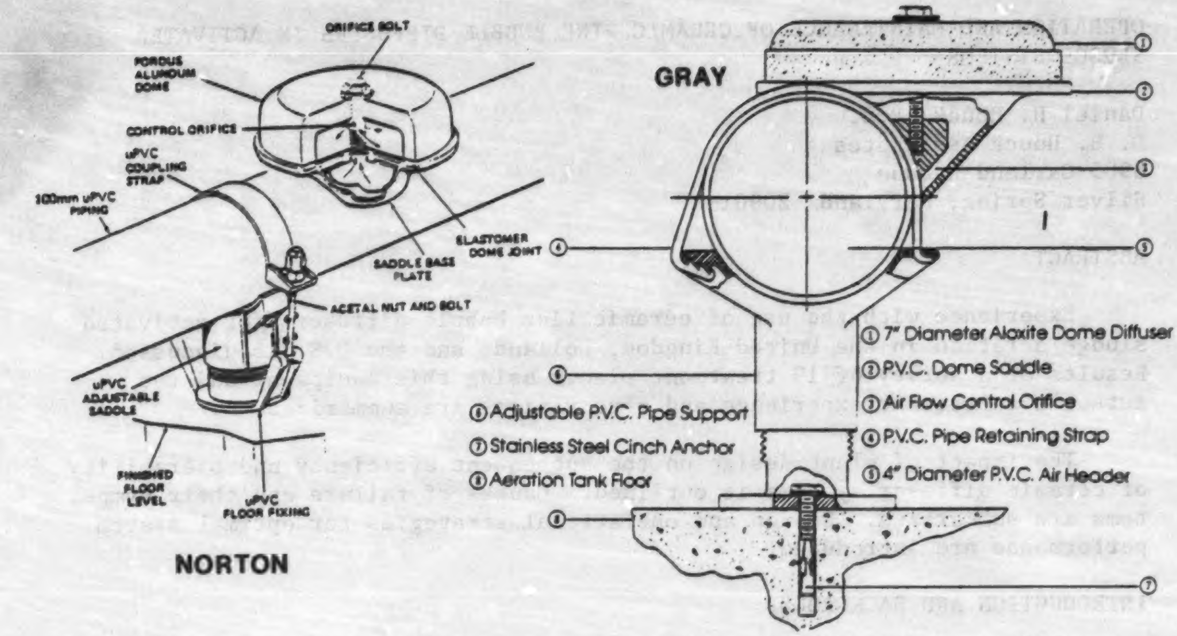

ORIFICE BOLT
PORDUS ALUMINUM DOME
CONTROL ORIFICE
uPVC COUPLING STRAP
100mm uPVC PIPING
ELASTOMER DOME JOINT
SADDLE BASE PLATE
ACETAL NUT AND BOLT
uPVC ADJUSTABLE SADDLE
FINISHED FLOOR LEVEL
uPVC FLOOR FIXING
NORTON
GRAY
①
②
③
④
⑤
⑥
⑦
⑧
① 7" Diameter Alaxite Dome Diffuser
② P.V.C. Dome Saddle
③ Air Flow Control Orifice
④ P.V.C. Pipe Retaining Strap
⑤ 4" Diameter P.V.C. Air Header
⑥ Adjustable P.V.C. Pipe Support
⑦ Stainless Steel Cinch Anchor
⑧ Aeration Tank Floor

Figure 1. Norton/Hawker Siddeley and Gray Engineering 7 inch dome diffusers.

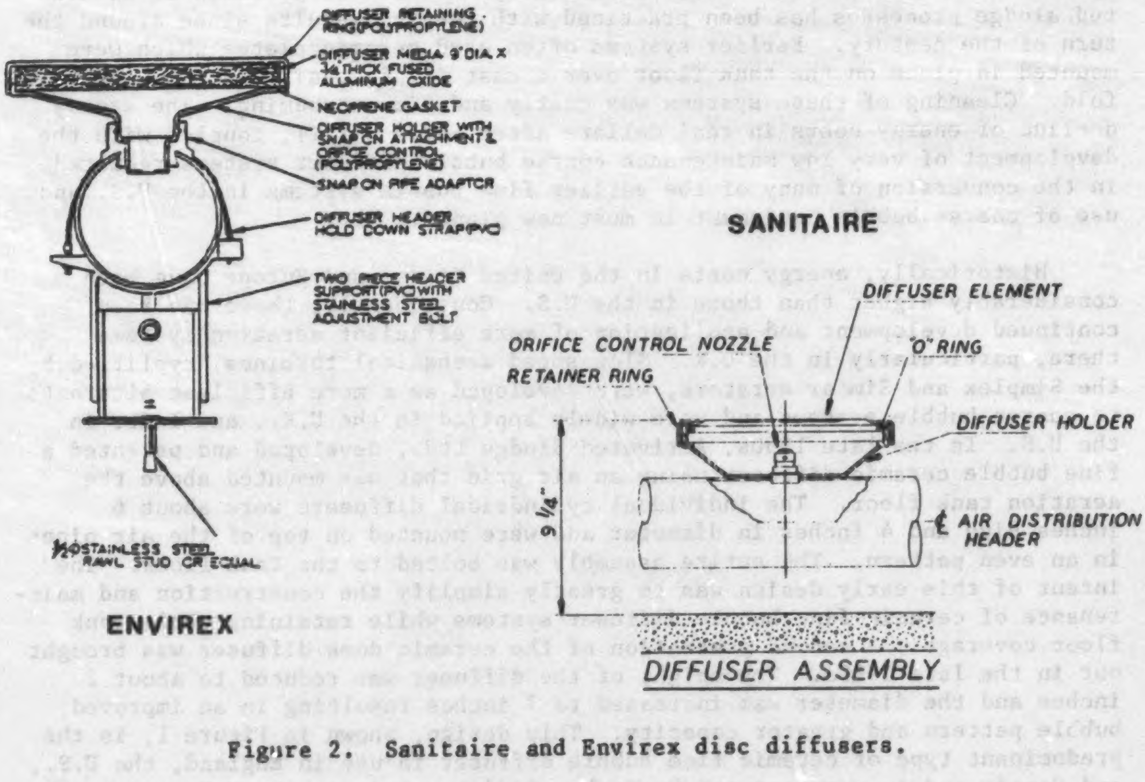

DIFFUSER RETAINING RING(POLYPROPYLENE)
DIFFUSER MEDIA 9" DIA. X 3/4" THICK FUSED ALLUMINUM OXIDE
NEOPRENE GASKET
DIFFUSER HOLDER WITH SNAP ON ATTACHMENT & ORIFICE CONTROL (POLYPROPYLENE)
SNAP ON PIPE ADAPTOR
DIFFUSER HEADER HOLD DOWN STRAP(PVC)
TWO PIECE HEADER SUPPORT(PVC) WITH STAINLESS STEEL ADJUSTMENT BOLT
1/2" ⌀ STAINLESS STEEL RAWL STUD OR EQUAL
ENVIREX
SANITAIRE
DIFFUSER ELEMENT
ORIFICE CONTROL NOZZLE
"O" RING
RETAINER RING
DIFFUSER HOLDER
9 1/4"
℄ AIR DISTRIBUTION HEADER
DIFFUSER ASSEMBLY

Figure 2. Sanitaire and Envirex disc diffusers.

Norton Co. in the U.S. Gray Engineering, a Canadian firm markets a similar 7 inch dome diffuser which features an offset dome saddle which they claim reduces the dome bolt stress on the pipe. Another type of ceramic fine bubble aerator uses a disc configuration. An example of this, shown in Figure 2, is manufactured by the Sanitaire Co. Disc diffusers are also manufactured in the U.S. by Envirex, Infilco-Degremont, and EPI-Nokia.

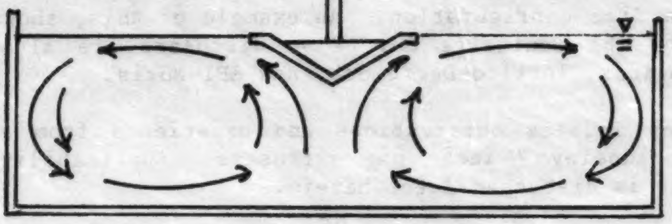
This paper relates observations and experience from systems using the Norton/Hawker Siddeley 7 inch dome diffusers. Applicability of the data to disc type units is discussed later herein.

## BASIC DESIGN CONCEPTS FOR DOME DIFFUSER ACTIVATED SLUDGE SYSTEMS

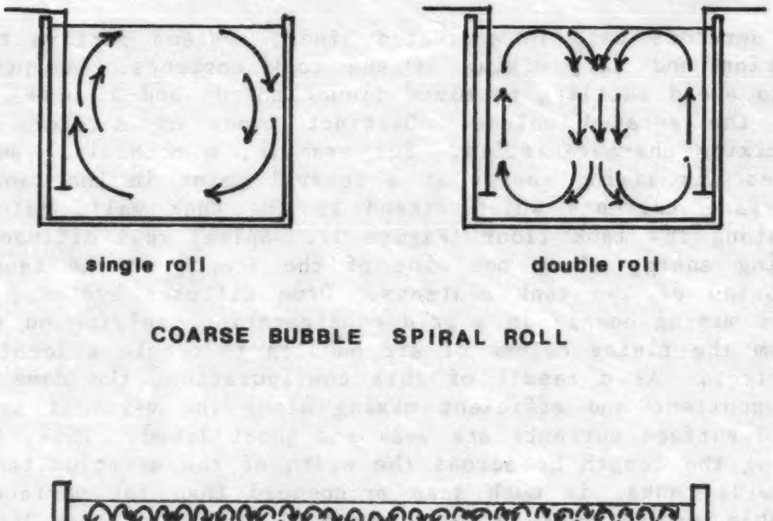
Operation and maintenance of dome diffuser activated sludge systems (DDAS) are very much affected by aeration system design. Inlet design and tank geometry in particular play a very large role in the operational characteristics of the system. Also, the system design should provide the operator with the means to monitor dissolved oxygen levels and control air flows to portions of the aeration tank if he is to have some control over the process.

Most aerators used in activated sludge systems perform two functions: (1), aeration and (2), mixing of the tank contents. Adequate mixing is required to avoid settling of mixed liquor solids and disperse the dissolved oxygen in the aerated volume. Distinct types of aerators have equally distinct mixing characteristics. For example, a mechanical surface turbine concentrates its mixing energy at a central point in the tank and creates strong surface currents which extend to the tank wall, returning to the impeller along the tank floor (Figure 3). Spiral roll diffused air systems input mixing energy along one side of the length of the tank, creating a rolling motion of the tank contents. Dome diffuser systems, on the other hand, input mixing energy in a grid configuration, relying on the release of energy from the rising column of air bubbles to create a localized air lift pumping effect. As a result of this configuration, the dome diffuser grid provides excellent and efficient mixing along the vertical axis. However, the induced surface currents are weak and short lived. Thus, the horizontal mixing along the length or across the width of the aeration tank, especially for very wide tanks, is much less pronounced than for surface turbines or coarse bubble spiral roll systems. Optimal design of dome diffuser aerated systems requires that the tank and inlet geometry take this characteristic into account.

Influent distribution configurations that work satisfactorily for coarse bubble aerators may not be adequate for ceramic diffusers. For example, influent was observed to channel along one side of a step feed tank that uses a series of gates in the tank wall[1], a common configuration in U.S. plants. This situation should be avoided in a new plant and extra diffusers and/or baffles should be placed around the inlet zones where existing plants are being converted to dome diffusers. Short circuiting has also been observed in the U.K. in the past and most systems there now combine the return sludge and settled sewage in a mixing channel prior to distributing it evenly across the width of the tank at the inlet end.

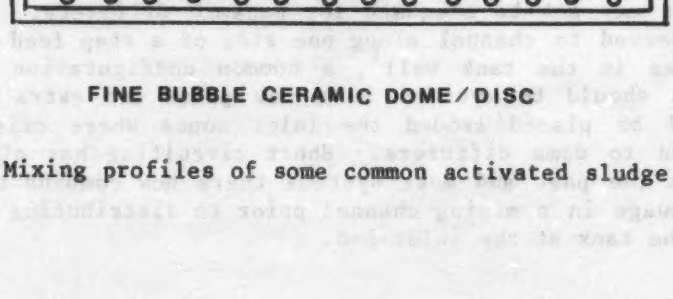


SLOW SPEED TURBINE



single roll double roll

COARSE BUBBLE SPIRAL ROLL



FINE BUBBLE CERAMIC DOME/DISC

Figure 3. Mixing profiles of some common activated sludge aerators.

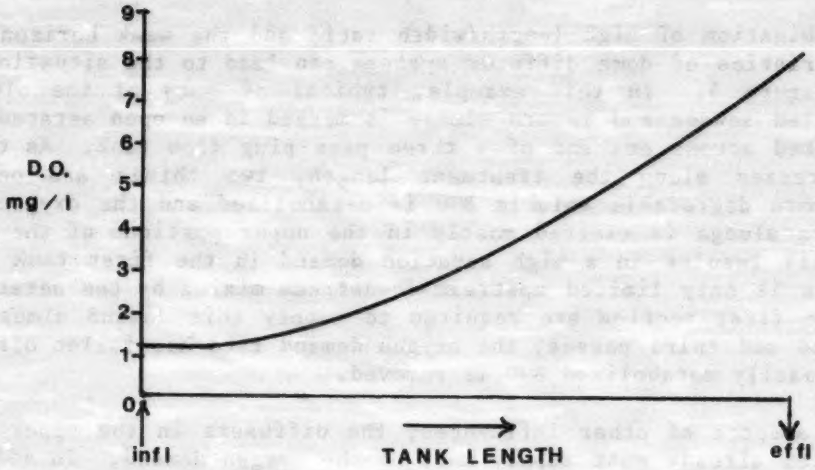


Figure 4. Rising dissolved oxygen profile in long, narrow plug flow tank.

A second problem, overaeration, was observed in many of the U.K. treatment plants visited by the author in an earlier study[2]. Excessive aeration of all or most of the aeration volume can result in poor energy efficiency of the system and may also adversely affect sludge settling. Plants that have this problem typically exhibit the dissolved oxygen curve shown in Figure 4. D.O. levels, in the proper range at the head of the tank, rise steadily along the tank length until they are quite elevated at the effluent end. This effect was repeatedly observed in U.K. plants and was most severe in plug flow systems with very high (over 20) ratios of tank length to width[2]. For the most part, the plant operators could not correct for the problem by adjusting air flows. Operators at the main plant for the City of London, Beckton, have removed up to one third of the aerators from some of the tanks in an attempt to reduce the overaeration effect. In doing so, they have verified that the system will maintain solids suspension at very low air flow rates per unit volume.

A variation of the overaeration problem occurs in the system that is simply too conservative in design. Such a system may have considerably more domes than required to meet the oxygen demand of the system. Each dome diffuser, and hence by extension the entire system, has a minimum air flow rate that it can be operated at. This minimum, 0.4-0.5 cfm for 7 inch domes, is required to avoid backflow of mixed liquor into the diffuser. Backflow can cause bio-fouling if the system is operated at lower air flow rates for extended periods. Seriously overdomed plants will supply excessive amounts of oxygen to the mixed liquor during much or even most of the daily treatment cycle. An extreme example of this was observed by the author at Strongford, England, where plant operators must run the system at minimum air flow for 20-22 hours per day and dissolved oxygen can range from 4-5 mg/l at the influent end of the tanks to near saturation at the effluent. Not surprisingly, the energy efficiency of this system is low.

The combination of high length/width ratio and the weak horizontal mixing characteristics of dome diffuser systems can lead to the situation illustrated in Figure 5. In this example, typical of many of the older U.K. plants, settled sewage and return sludge is merged in an open aerated channel and distributed across one end of a three pass plug flow tank. As the mixed liquor progresses along the treatment length, two things are occurring. First, the more degradable soluble BOD is metabolized and the oxygen deficit of the return sludge is exerted mostly in the upper portions of the aeration process. This results in a high aeration demand in the first tank section. Because there is only limited upstream-downstream mixing by the aerators, the domes in the first section are required to supply this demand almost alone. In the second and third passes, the oxygen demand rate has fallen off sharply as the less easily metabolized BOD is removed.

In the absence of other influences, the diffusers in the upper sections of this system already must supply most of the oxygen demand. In addition, a second condition makes the load on the upper tank sections even more extreme. The alpha factor in this situation is also variable! The projected curve shown in Figure 6, from observations in other similar situations[3] indicates a 100 percent variation in alpha from front to back of the three pass system may be possible. In concrete terms, a dome diffuser in the first part of the first pass is only one half as efficient as one located immediately before the outlet. At the same time, the first dome is operating in thea zone of highest oxygen demand, where D.O. requirements may be 2-3 times higher than in the lower zones of the system.

Tapering the aeration, placing more of the diffusers in the first sections of the tank, has been recommended for quite some time by both the U.K. and U.S. manufacturers of dome diffusers. However, this remedy only works up to the point where mixing requirements begin to control the number of diffusers required in the lower portions of the aeration tank. Beyond that point, any more diffusers needed in the upper aeration zones must be added to the total, and overaeration in the lower zones is likely to result. It is the author's opinion at this point that tanks with length/width ratios exceeding 16/1 cannot reap the full benefit of the efficiency of ceramic dome or disc aerators due to geometric limitations.

Many of the plants visited by the author in an earlier study[2] suffered from combinations of poor tank geometry and maldistribution of aeration. Even where aeration was tapered, tapering was inadequate in situations where length/width exceeded 10/1. Thus, an operator confronted with a plant designed like this can do little to optimize its performance. He is lucky if he can avoid slime growth and other operational problems which result from poor tank/inlet configuration.

If the designer does not provide a reliable dissolved oxygen monitoring system and adequate air valving and controls, the operator will have little incentive for and great difficulty with optimization of D.O. levels. D.O. levels are most critical in the first section, where very low oxygen can lead to slime growth on diffusers and piping, and in the final section where there may be a tendency to over aerate. At minimum, D.O. should be monitored and air flows individually controlled at these points. Other control points may

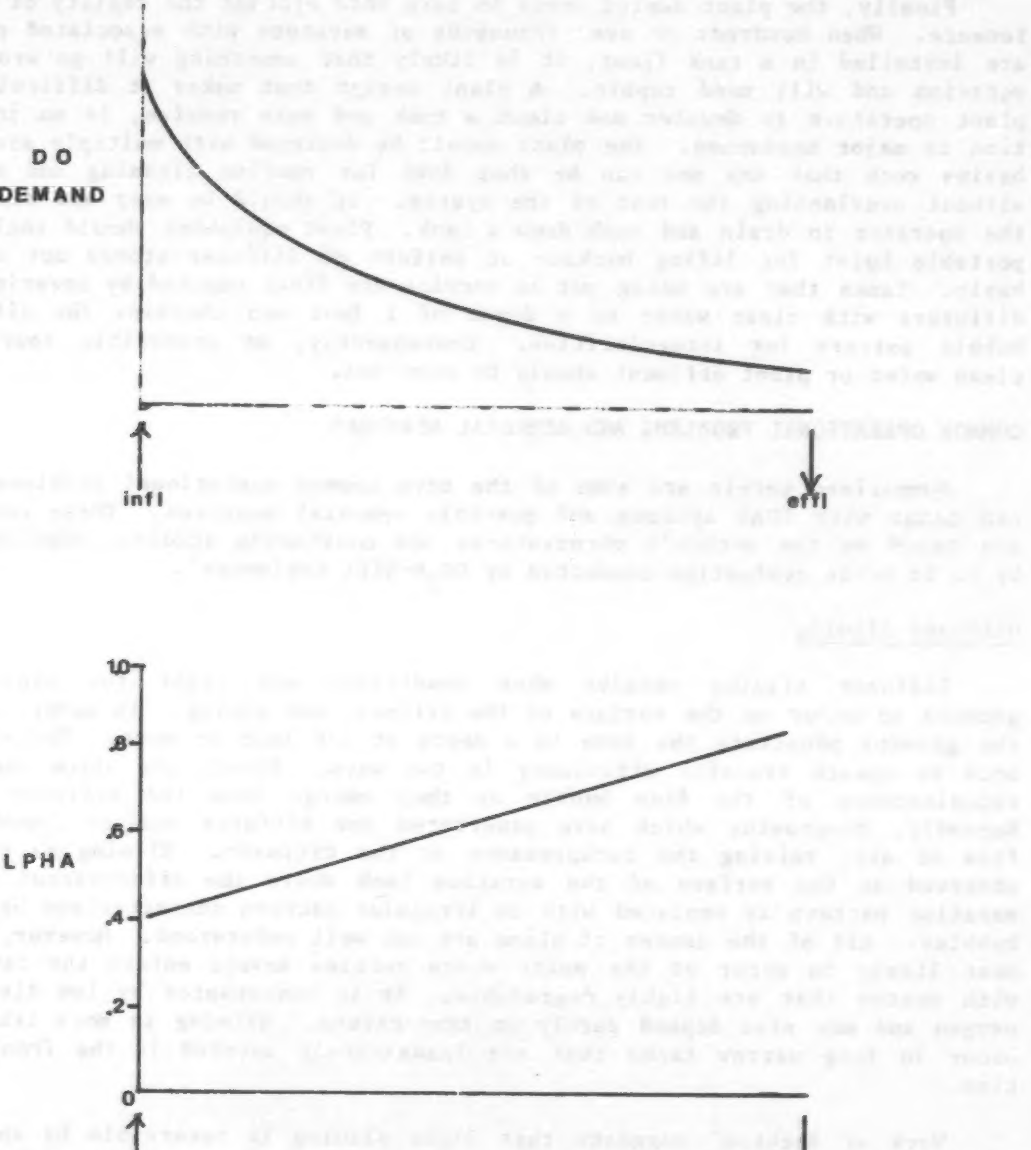


Figure 5. Oxygen demand profile and alpha variation in extreme plug flow systems.

be required depending on system geometry.

Finally, the plant design needs to take into account the reality of maintenance. When hundreds or even thousands of aerators with associated piping are installed in a tank floor, it is likely that something will go wrong on occasion and will need repair. A plant design that makes it difficult for plant operators to dewater and clean a tank and make repairs, is an invitation to major headaches. The plant should be designed with multiple aeration basins such that any one can be shut down for routine cleaning and repair without overloading the rest of the system. It should be easy and safe for the operator to drain and wash down a tank. Plant equipment should include a portable hoist for lifing buckets or pallets of diffuser stones out of the basin. Tanks that are being put in service are first checked by covering the diffusers with clear water to a depth of 1 foot and checking the diffuser bubble pattern for irregularities. Consequently, an accessible source of clean water or plant effluent should be provided.

## COMMON OPERATIONAL PROBLEMS AND REMEDIAL MEASURES

Summarized herein are some of the more common operational problems that can occur with DDAS systems and possible remedial measures. These comments are based on the author's observations and continuing studies, supplemented by an in house evaluation conducted by $CH_2M$-Hill Engineers[1].

### Diffuser Sliming

Diffuser sliming results when conditions are right for biological growths to occur on the surface of the diffuser and piping. In severe cases, the growths penetrate the dome to a depth of 1/8 inch or more. The sliming acts to reduce transfer efficiency in two ways. First, the slime causes a recoalescence of the fine bubble as they emerge from the diffuser face. Secondly, biogrowths which have penetrated the diffuser surface impede the flow of air, raising the backpressure at the diffuser. Sliming is readily observed at the surface of the aeration tank where the effervescent gentle aeration pattern is replaced with an irregular pattern characterized by large bubbles. All of the causes of slime are not well understood. However, it is most likely to occur at the point where settled sewage enters the tank and with wastes that are highly degradable. It is exacerbated by low dissolved oxygen and may also depend partly on temperature. Sliming is more likely to occur in long narrow tanks that are inadequately aerated in the front section.

Work at Beckton[2] suggests that light sliming is reversible by shutting off the flow of untreated waste to a tank and aerating recirulated flow only for 24-48 hours. More severe cases will require draining of the tank and cleaning of the diffusers. At Beddington[2], the diffusers are periodically brushed to remove growths. Madison[1] steam cleaned badly slimed diffusers, which seemed to cure the problem, although the pressure loss through the diffusers remained somewhat elevated after cleaning. Badly fouled diffusers may require refiring in a kiln, which necessitates their removal and replacement. This approach, commonly used in the U.K., costs about one half of the cost of new diffusers. At this time, there is no one method that is wholly satisfactory for dealing with sliming. The Sanitaire Corp. has been developing an in-line acid cleaning system, which is currently being installed in several plants. However, there is inadequate experience with the system at this point in time to conclude that it will be able to deal with diffuser fouling, particularly of the organic variety. At this point in time, the most cost effective means of dealing with diffuser fouling is to prevent it through proper design and operation.

## Diffuser Plugging

Diffusers can also be plugged internally by dirt in the air supply, or by rusting of metallic pipes. In either case, the material is deposited on the inside of the dome, gradually plugging up the pores of the ceramic material. In a few cases, inorganic scaling on the exterior surfaces of domes has occurred[2]. Acid washing has been used to remove both types of plugging. The best remedy for internal fouling is to avoid the problem by maintaining air cleaners properly and by using non-metallic pipe materials where ever moisture may be present, particularly below the water lines in aeration tanks. Scaling may not be preventable and use of ceramic fine bubble diffusers should probably be avoided where this is known to be a severe problem.

## Failure of Plastic Pipe and Fittings

Out of 6 plants visited by the CH$_2$M-Hill team, problems were experienced at 5 with plastic piping and hardware.[1] The sixth plant, Toronto, uses metal piping and has had no failure with the exception of possible leakage around the dome gasket. The most common failure reported is breakage of the plastic dome retaining bolt. This parallels the British experience and is usually caused by overtightening of the bolt by installers. Other causes of failure include cracks in the pipe saddle, failure in expansion joints, blow off line breaks, and air leakage around the dome gasket. Overtightening was cited as the principal cause of most of the failures at the plants visited by CH$_2$M-Hill. There were indications at one plant that assembly of the piping during hot weather may have caused failures as well. Figure 6, from Reference 1, illustrates the principal failure modes.

Interestingly, the implied frequency of failure in the CH$_2$M-Hill work may be somewhat greater than that experienced in the U.K. Possibly this is due to the fact that most of the aeration grids in the U.K. were installed by the equipment manufacturer, Hawker Siddeley, with experienced crews. Plant operators in the U.K. note that proper initial installation is of primary importance for a relatively troublefree installation. Also, British operators expect and accept the need to periodically maintain dome diffuser systems. Dewatering of tanks, cleaning and checking are routinely performed every 2-3 years in most U.K. plants. With routine maintenance of this type, there is a greater liklihood that problems will be corrected before they become serious.

## APPLICABILITY OF DOME DIFFUSER FINDINGS TO DISC SYSTEMS

As noted previously, several U.S. and foreign manufacturers have developed disc aerators (Figure 2) which are somewhat similar to dome diffusers.

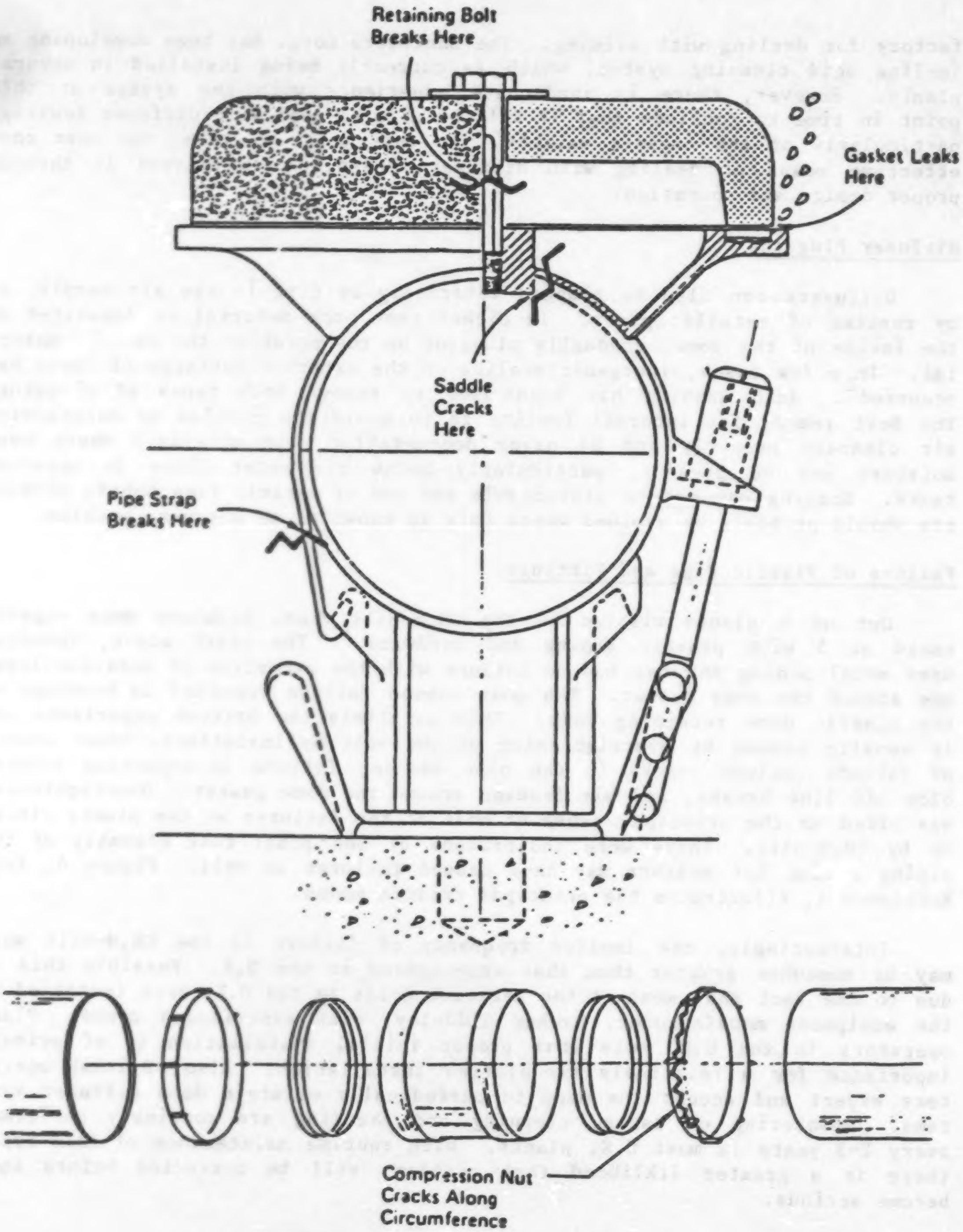

Retaining Bolt
Breaks Here
Gasket Leaks
Here
Saddle
Cracks
Here
Pipe Strap
Breaks Here
Compression Nut
Cracks Along
Circumference

Figure 6. Dome diffuser and plastic piping failure modes[1]

Often, these units are being bid as alternates on projects specified for dome diffusers. Several efforts[1,2,4] have been made to assess the substitutability of disc diffusers for domes. Experiments at Holten-Markelo, in Holland, suggested that oxygen transfer between dissimilar sized ceramic diffusers is directly proportional to surface area.[2] More extensive work at Los Angeles County, where the 8.7 inch Sanitaire disc was compared to the 7 inch Norton dome, concluded that 3 of the discs would transfer about the same amount of oxygen as 4 domes[4]. However, mixing has not been addressed specifically in any of these studies. Substitutions of this type should be reviewed closely before being approved as they might lead to mixing problems in a small number of cases where diffuser density and air flow are already very low.

Mechanically, the disc diffusers may offer several advantages over dome type units. The elimination of the center bolt eliminates the problems caused by bolt stress on the pipe saddle (Gray Engineering's offset design also mitigates this problem). Applying hold down pressure uniformly around the periphery of the diffuser is also more likely to prevent leaks around the base gasket. However, to date experience with these units is rather short and long term problems may yet occur.

## CONTINUING RESEARCH

Spurred by the U.S. EPA, the British Water Research Centre, and Environment Canada, major research efforts in the last 4 years have yielded significant strides in understanding ceramic diffuser system design and operation. Current research is focusing on optimization of the efficiency of these systems and in avoiding or solving problem areas. Substantial changes in past design and operational practices is already underway and current and future research should result in further advances. Presently, there are three major research projects underway:

The Reymeads, U.K., Demonstration Project- Based on the initial investigations detailed in Reference 2, and further developmental work by the WRC, a full scale demonstration of optimized design and opeation of a dome diffuser system has been set up at the Ryemeads plant. One third of the plant, treating 3.2 mgd, has been revamped to improve the diffuser configuration and provide dissolved oxygen monitoring and automatic air flow control. Substantial increases in measured system efficiency have already been achieved, and further modifications were made recently which should result in overall efficiency improvements in excess of 100 percent over the existing system.

Comparative Studies of Fine Bubble Aeration Systems, Los Angeles County, California- A joint project between the U.S. EPA and Los Angeles County is conducting detailed efficiency and design studies of a number of fine bubble aeration systems, including domes, discs, jet aerators, and tubular units.

Biofouling/Performance Studies at Madison, Wisconsin- Efficiency measurement using off-gas analysis is the focus of this project, which is also studying the causes and cures of the episode of bio-fouling experienced at this plant in 1980.

From the results of these and other work, a design rationale for plug and step feed applications of ceramic grid fine bubble diffuser systems is currently being developed by the author of this paper. Using a micro-computer to handle tedious calculations, the design model treats the aeration tank as a series of completely mixed square reactors and considers the variation in alpha and oxygen demand with detention time (tank length). When finished, the program will be used to calculate the number of diffusers required at any point in the tank(s) with much greater precision than current practices. Of course, the output from this design model must be coupled with experience gained from study of existing systems in arriving at the final design.

## CONCLUSION

Extensive experience in the United Kingdom with ceramic dome diffusers has shown them to be efficient and reliable when applied and operated properly. The key to effective use of this technology is understanding its design and operational boundaries. Aeration basin and inlet geometry is of prime importance in new or retrofitted plant design. Operators of these systems should be trained to spot the symptoms of impending problems and respond to them quickly. Close dissolved oxygen monitoring and control is a must for system efficiency and control of bio-fouling. Maintenance need not be excessive, but it must be routine and thorough for best results. Ceramic diffuser grid systems are not "install it and forget it" systems! Even so, the cost incurred for additional maintenance is offset many times by the 200-300 percent improvements in electrical efficiency that can be realized over coarse bubble systems. Other advantages include lack of noise and airborne aerosols, and ease of plant sanitation. The savings in power costs, coupled with other operational advantages, have resulted in fine bubble ceramic grid systems becoming the system of choice in the U.K. North American plant designers and operators can expect the same benefits from this technology if they are willing to apply it correctly and benefit from the more than 20 years of overseas experience with ceramic fine bubble aeration systems.

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## DISCLAIMER

The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency, and no official endorsement should be inferred.

# OPERATION AND MAINTENANCE OF AERATION SYSTEMS - AN OPERATOR'S PERSPECTIVE

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## ABSTRACT

This paper will deal with the most commonly found problems encountered in the operation and maintenance of aeration systems in Wastewater Treatment Plants. The observations will be made from the Operator/Troubleshooter perspective and will be based on the author's 16 years of experience in troubleshooting and operating Pollution Control Plants. Each generic type of equipment will be discussed and both positive and negative aspects noted.

## MECHANICAL SURFACE AERATORS

There are currently a wide variety of mechanical surface aerators available for the pollution control market. They may be fixed mounted or floating, low speed or high speed, rotate in the vertical or horizontal plane, be gear driven, chain driven or belt driven, have individual drive motors or line shaft driven, with or without draft tubes, and even have lower mixing impellers.

This extreme variability makes it difficult to generalize on the type of problems associated with mechanical surface aerators. However, there are a number of reoccurring problems that warrent discussion.

The inability of surface aerators to transfer the required or designed oxygen capacity can frequently be traced back to the submergence sensitivity of these devices. Both the power draw and the oxygenation capacity are directly related to the submergence of the aeration device into the liquid. Although some proprietary devices claim to be less submergence sensitive, all share the same tendency. Frequently changes in flow and/or downstream level control devices leave the surface aerators with inadequate submergence and a resultant oxygen deficiency.

The loss of buoyancy through the accumulation of ice or the waterlogging of floatation material produces the opposite effect of excessive submergence in floating aerators with the resultant power overload and inadequate oxygenation capacity.

Alteration of rotational speed through gearing changes or motor changes can frequently lead to loss of aeration capacity. Less frequently, changes to the aerator blades imparts a negative impact.

Mechanical problems with the gearboxes or motors were at one time a significant problem. In the last decade these problem areas have been redesigned resulting in long term trouble free operation from much of the equipment being sold today. As a result the problem of maintaining a large number (20 to 50) mechanical aerators at a large plant are not as onerous as once was the case.

Mechanical aerators have several inherent advantages, namely: relative-high oxygenation efficiency, a great deal of operational flexibility, low impact of mechanical failure (i.e. if one of 20 aerators is out of service the impact on treatment efficiency is low), generally can be used in existing shallow tanks.

The effect of wastewater characteristics such as low alpha values frequently have less of a negative impact with mechanical aerators than in many diffused air devices.

## DIFFUSED AIR

The use of diffused air aeration systems has been common for most of this century in Water Pollution Control Plants. There are two basic system types available on the market, namely: course bubble and fine bubble systems.

### Fine Bubble Diffusers

Historically fine bubble systems were plaqued with airside clogging problems associated with air borne dirt. No satisfactory cleaning systems were developed until in the last decade when new materials of construction, such as PVC pipes, for air distribution and improved low cost air filtration systems were introduced to the market.

The advent of this new technology along with the need to improve the energy efficiency of diffused air systems has led to a proliferation of fine bubble air diffusion systems reaching the market. Many of the systems on the market today have not fully realized the mechanical difficulties of operating a diffused air system in an aeration tank. The equipment tends to be far to fragile and subject to corrosion leading to premature mechanical failure. It is imperative that the equipment designers appreciate the need for long term ($10^+$ years) mechanical reliability and allow for this in their design. There is nothing more frustrating then needing to dewater aeration tanks to fix a substandard piece of aeration equipment and then having to redo it six months later because the same component has failed again.

The solving of the air side plugging of fine bubble aeration devices has not, however, totally solved the fouling problems. There still exists a significant potential of wet side plugging associated with either biological slime growth or deposition of inorganic solids.

The growth of biological slimes on the surface of fine bubble devices is usually related to low oxygen concentrations and/or high organic loads. Both of these causes can be handled adequately in the design of the aeration system by ensuring adequate DO levels are maintained at the head end of the aeration tank where the organic load and resultant oxygen demand are highest. The demands can be met by providing a tapered aeration system or step feeding the organic load down the length of the aeration tank.Both solutions have been found to rectify problems with biological fouling.

The blinding of fine bubble devices by the deposition of inorganic solids, generally considered to be calcium carbonate, has continued to be a problem in the area of industrial waste treatment where the dissolved solids are high. The potential for inorganic blinding increases in direct proportion to the level of dissolved inorganics.

In municipal sewage treatment inorganic fouling has not commonly been a problem and would not be anticipated unless the municipal water supply was extremely high ($500^+$ mg/L) in hardness, or significant industrial inputs raised the raw sewage dissolved solids to the 1500 mg/L range.

A recent inspection of a one year old fine bubble aeration system installed in a high rate activated sludge plant (200 lb BOD/1000 cu ft/day) treating tannery wastes showed almost complete inorganic fouling of the dome diffusers. The ceramic diffusers were coated with a heavy layer up to 1/8" thick of a smooth gray inorganic deposit that was readily dissolved in dilute hydrochloric acid, generating $CO_2$ gas.

The small areas not covered by the inorganic solids were coated with biological slime. During the year of operation the air pressures increased by only 1.0 to 1.5 psi. Inspection showed that almost all the air was escaping at the seat gasket in a coarse bubble pattern. No evidence of air side fouling was noted.

The domes were restored to a visually clean situation by immersing them in dilute hydrochloric acid overnight and then rinsing them with water.

The tannery wastewater has a dissolved solids content of 8000 mg/L with a major contributor being lime used in the dehairing process. It is anticipated that inorganic fouling will reoccur if some means of in situ cleaning is not installed. An H Cl gas cleaning system is currently under design complete with a head loss monitoring system to ensure that remedial action is taken before total clogging occurs. It is anticipated that perhaps even weekly cleaning may be required.

Mechanical inspection of the aeration system showed that several of the support brackets had come loose from the floor by unscrewing the holddown bolt, likely due to system vibration.

No pipe, dome, or bolt breakage was evident.

An adjacent equalization tank, agitated using coarse bubble diffusers, showed significant inorganic deposition on the sparger type diffusers but not sufficient to cause plugging of the air flow. A significant number of diffusers were plugged with organic debris associated with the tannery wastewater.

## Aeration Blowers

The operation of small (up to 150 HP) positive displacement blowers particularly when operating at rotational speeds of 1000 rpm or less has proven to be very reliable. It is not uncommon to experience ten or more years of trouble free operation with only normal lubrication schedules being followed. As the rotational speed increases the rate of wear and the noise level increase dramatically, resulting in more frequent breakdowns.

The use of multi stage centrifugal blowers particularly those operating at 3600 rpm has proven to be less satisfactory. In some cases major bearing and/or vane problems have been experienced yearly while other facilities appear to have long term trouble free operation. It is apparent that the higher rotational speeds used in these blowers demand far closer alignment tolerances than seen in the lower speed positive displacement blowers. The results of any misalignment are premature motor bearing failure; the most common operational problem with centrifugal blowers.

In large plants the use of a small number of large centrifugal blowers frequently leads to a severe lack of flexibility in the air supply system, even when using the limited turn down capacity of these systems. Frequent overaeration during low flow periods particularly before the plant reaches its design load is a common problem. This does not reflect badly on effluent quality but it does result in a significant loss of energy efficiency. The use of a large number of smaller blowers to allow the air supply to match the organic load would be preferable.

## Coarse Bubble Diffusers

The mechanical and process reliability of coarse bubble diffusion devices make them ideally suited for use in activated sludge systems. Many coarse bubble systems have been in service for 5 to 10 years without being inspected or causing operational problems. When used in municipal wastewater seldom is clogging a significant problem. Some industrial wastes containing stringy organic solids such as hair or fleshings may cause plugging of the orfice holes.

The major drawback to the use of coarse bubble diffusers is their low energy efficiency. A wide variety of changes in header and diffuser spacing as well as deep tank configurations have made minor improvements to the energy efficiency but has not been able to approach the efficiency seen in fine bubble devices or mechanical aeration systems.

## Hybrid Aeration Devices

In attempts to improve aeration electrical efficiency a large number of hybrid aeration devices, combining air blower or aspirators with either mechanical mixers or pumps, have been marketed. Some are intended to be operated as surface devices while others are floor mounted subsurface systems. A recent innovation of a rotating subsurface diffusion system also falls into this category.

From an operator's perspective these systems, although offering some improvement in efficiency, also result in increased potential for mechanical failure. Experience has shown that annual pump and/or blower failures are seen with nozzle plugging not being uncommon particularly in systems carrying high (5000$^+$ mg/L) aeration solids.

The use of these devices in long detention systems to prevent winter icing problems on mechanical aerators has been satisfactory. Parallel operation of two long detention (5$^+$ days) activated sludge systems treating brewery wastes in central Ontario, one utilizing surface aerators and the second using a jet aeration system, have shown a dramatic difference in winter aeration tank temperatures. The mechanically aerated system operated at $-1^\circ$C with almost a total ice cover on the aeration tank. The jet aeration system operated at $4^\circ$ C with little to no ice cover. Both systems were fed equal amounts of wastewater and are physically identical in size and depth. The advantage of reduced heat loss particularly in northern climates is obvious from an operational viewpoint as well as providing increased levels of biological activity.

This advantage of reduced heat loss is shared by all subsurface aeration devices and is used to advantage in aerated lagoons by the use of static mixer type aeration systems. There are a number of generic devices on the market designed to be used in aerated lagoons to provide both mixing and oxygen supply. These devices share the principal that the air bubbles and the moving water column are confined in a vertical tube and mixed by some physical mixing system. The aeration efficiency is improved above straight coarse bubble aeration.

Problems with poor anchoring system and plugging of the mixing devices are common. The use of these aerators in activated sludge plants has proven to be less than satisfactory because of their inability to maintain the aeration solids in suspension at normal aerator spacings. If the aerators are brought closer together to improve the mixing regime the aeration efficiency is significantly reduced.

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# OPERATION AND MAINTENANCE/TROUBLESHOOTING

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The subject matter of this paper deals primarily with areas affecting the operation and maintenance of fine bubble aeration systems, particularly grid systems employing domes, discs and plates. The above areas, however, are applicable to fine bubble diffusion systems in general.

Due to the relatively high oxygen transfer efficiency of fine bubble, ceramic grid equipment, many engineers are considering these systems for retrofits as well as new wastewater treatment facilities. Considering recent studies that indicate 50 to 90 percent of the net power demand for a treatment plant lies within the aeration system (1), it is not surprising that energy efficient aeration systems are being sought out.

Various clearwater data of dome and disc fine bubble systems (2,3,4), certainly substantiate the high clearwater efficiencies, and recent offgas analyses of these systems indicate very efficient dirty water performance as well, when compared to other diffused systems studied (5). A discussion of the offgas technique and a highlight of the results will be given in Session 4 of this seminar.

Concern, however, has been registered by some regarding the maintenance of fine bubble systems owing at least partly, to their suspectibility to clogging and the mechanical integrity of the system.

Houck and Boon (6), who carried out an extensive survey of 19 wastewater treatment plants, employing dome diffuser systems, in the United Kingdom, Netherlands, and the U.S., indicated the need for system maintenance. The major operational problem reported was associated with the formation of biological slime on diffusers operating in zones of high biological loading and/or low dissolved oxygen.

In January of 1982, Boyle and Redmon (7), presented a paper entitled, "Biological Fouling of Fine Bubble Diffusers", which brings together much of the information on biological fouling.

Fouling may lead to a deterioration in efficiency both from a standpoint of mass transfer and elevated pressure to operate the system. In addition, fouling may incur substantial costs for cleaning and maintenance.

The economic impetus to operate these systems in an optimal fashion both from an efficiency and maintenance viewpoint justifies careful specification of the equipment, system survelance, and fouling diagnosis of the diffusion system, on an on-going basis. The latter can be used to evaluate the degree of fouling and system performance with time and can be used as a basis in the optimization of cleaning cycles.

## MECHANICAL ASPECTS

In attempting to minimize operational problems and maximize dirty water efficiency, attention should be paid to the mechanical aspects of the equipment. One should strive to achieve good distribution of airflow among the units on a given grid through the use of properly sized individual control orifices. Since the wet pressure drop across the ceramic itself is quite insensitive to flow rate, balancing is required to assure reasonable distribution among the elements. Poor distribution of air greatly reduces the efficiency and may exacerbate the fouling problems.

Regarding systems piped above the tank bottom, such as Norton, Sanitaire, Rexnord, Carbo Grey, etc., the support system should be capable of withstanding uplift, thrust due to expansion and contraction forces, as well as loss of air pressure, and should be capable of simple elevation adjustment to within ± 1/4 inch. The expansion joints should operate with minimal forces with or without dependence on lubricants and be able to provide leak-free operation at a positive internal pressure differential of up to 4-5 psi, while sustaining shear loads and/or bending forces which the system will be subject to in operation.

For high temperature applications, special care should be taken regarding the long term stability of the plastic systems. At elevated temperature and/or stress, plastic systems may represent a problem if not properly engineered to operate in these environments.

## DIFFUSER CHARACTERIZATION METHODS

In specifying a ceramic system proper specifications of the ceramic element itself is of great importance. Some areas of interest include the desired specific permeability of the element, it's wet pressure characteristics and the uniformity of air release across an individual ceramic element.

Ewing Engineering Company has been testing, developing and applying fine bubble diffusers over the past 4-5 years. During this period we have developed a series of test to quantitatively measure fine bubble diffuser characteristics on new and fouled diffusers.

These tests measure ceramic wet pressure drop versus flow rate, the uniformity of the surface pores and uniformity of air release. A general description of the parameters, specific permeability, dynamic wet pressure, bubble release pressure and air flow profile are presented below.

### Specific Permeability

The manufacturers of ceramic diffusers have used and are familiar with the permeability test. It has served as a quality control procedure to assure that the units sent to a jobsite are similar with respect to their average frictional resistance to flow, when dry, to within some specified limits. This was especially important in many older plants where several plates were installed into a single plenum without individual flow balancing means toward the objective of improved uniformity of air flow among the units when in operation.

The test generally consists of sealing the ceramic unit in a test fixture substantially, as it is sealed in an actual aeration tank, and then passing sufficient air through the dry element to produce a pressure differential of 2.0 inches water gauge. The permeability is reported as the airrate required to produce this differential. In the U.S., the airrate is in SCFM, (standard cubic feet per minute), where a standard cubic foot of air is considered to occupy 1.0 cubic foot of volume at one atmosphere, (14.696 psi; 760 mm HG), 70 $^{o}$ F, and 36% relative humidity. Historically, the test was carried out on ceramic plates 12 inches x 12 inches x 1 inch thick.

In the way of an example, if we ran a permeability test on such a ceramic plate, we might find it took 25 SCFM to produce a 2.0 inch wg. differential pressure. In this case, the permeability rating would be 25. A plate of identical material, but half as thick, would be expected to have a permeability of 50 instead of 25, since the flow paths through the ceramic would be about half as long and offering correspondingly less frictional resistance. Had the element been 1.0 inch thick, and had an area of 72 sq. inches instead of 144 sq. inches, the permeability would be approximately 12.5 since there would have been only about half the area of the first case.

Even though the above examples had identical pore size and configurations, the permeabilities of the three vary from 12.5 to 50. Thus, using permeability to compare ceramic elements of different shape, thickness, materials of construction and the like, is not meaningful and has been a confusing factor in the engineering community.

In an effort to employ permeability test results as a measure of resistance characteristics of the material, we adopted the term "specific permeability", which is the equivalent amount of air at standard conditions to produce a 2.0 inch differential pressure across the dry element if the element were 1.0 sq. foot in area, (12 inch x 12 inch), and 1.0 inch thick.

An approximate expression to convert the permeability of any porous structure to specific permeability is as follows:

$$S.P. = P \times \frac{t}{A}$$

Where: S.P. = specific permeability, SCFM

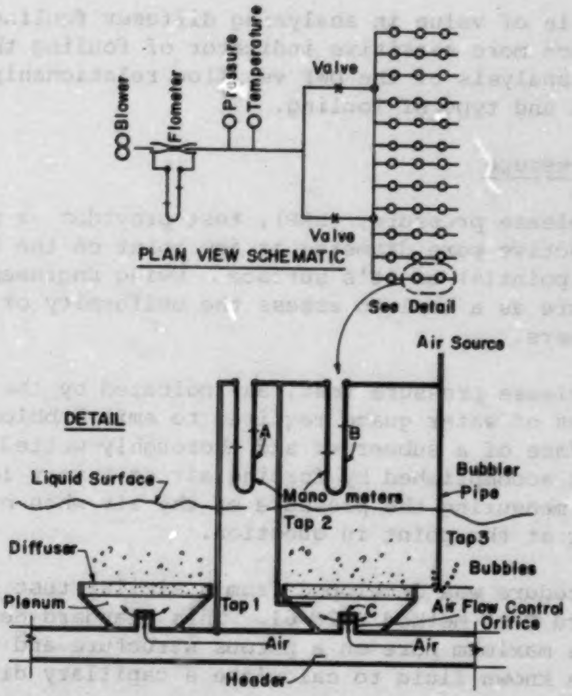
P = permeability of the element itself, SCFM

A = area of element, sq. ft. when made to hypothetically conform to a flat surface

t = mean weighted thickness of the element, inches

## Dynamic Wet Pressure

The dynamic wet pressure, DWP, is the pressure differential across the diffusion element alone when operating in a submerged condition and is expressed in terms of inches of water, at some specified flow rate. It differs substantially from the permeability test, where the specific air flow rates are typically much higher and where surface tension of the submerging liquid is not a factor.


Blower
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Bubbler Pipe
Tap 2
Tap3
Diffuser
Bubbles
Plenum
Tap 1
C
Air Flow Control Orifice
Air
Air
Header

Figure 1. Measurement of Air Line and Diffuser Pressure

In the dynamic wet pressure test, most of the pressure differential is due to the force or pressure required to form bubbles against the force of surface tension and only a small fraction of the total pressure gradient is required to overcome frictional resistance.

The DWP is normally relatively insensitive to airflow rate. This results since the surface tension effect is not greatly influenced by flow rate in flow ranges normally applied and the contribution of frictional resistance is a small part of the total pressure gradient.

That the frictional resistance is small with respect to the total, can be shown as follows:

> Assume the typical airflow rate at average design conditions is about 2 SCFM per square foot of diffuser surface area. Since the frictional resistance is only 2.0 in. wg. at 25 SCFM, (specific permeability of 25), and this unit is typically operated at 2 SCFM under average design conditions, it's theoretical frictional resistance at this rate is roughly 2.0/25.0 of 2 in. wg. or about 0.2 in. wg. The total DWP at 2.0 SCFM is about 6.9 in. wg. So the frictional resistance at this airflow rate is roughly 3% of the total resistance.

The DWP test is of value in analyzing diffuser fouling in at least two ways. It is a much more sensitive indicator of fouling than air main pressure readings and the analysis of the DWP vs. flow relationship gives some indication of the nature and type of fouling.

## Bubble Release Pressure

The bubble release pressure, (BRP), test provides a means of comparing the relative effective pore diameter at any point on the surface of a ceramic element to other point(s) on it's surface. Ewing Engineering Company developed this test procedure as a tool to assess the uniformity of pores on the surface of ceramic diffusers.

The bubble release pressure test, as indicated by the name, measures the pressure in inches of water guage required to emit bubbles from a localized point on the surface of a submerged and thoroughly wetted porous diffuser element. This is accomplished by forcing air at a very low rate of flow into the diffuser and measuring the pressure of the air when bubbles are released from the diffuser at the point in question.

The test procedure was developed from a similar test found in ASTM, Part 41, under Standard Test Method E128-61. This standard describes a technique for measuring the maximum pore on a porous structure and uses the force to make bubbles in a known fluid to calculate a capillary diameter corresponding to this force or pressure.

The hardware employed in this test has been modified for this application. BRP observations of a statistically significant number of points on the element were obtained to ascertain the degree of uniformity. After determining the air flow distribution on several elements on which BRP data was available, it became apparent that the BRP test data was a sensitive indicator of flow distribution.

This test has been applied to a variety of fine pore diffuser elements. It has been found to be a sensitive indicator of the degree of fouling of porous diffusers and provides a quantitative means of assessing the rate and location of plugging.

Furthermore, analysis of partially fouled diffusers have given indication that the BRP test is even more sensitive than the DWP test. Consequently, in diffuser fouling investigations, the use of these test methods permits significant shortening of the test period required to obtain definitive conclusions about diffuser clogging.

## Air Profile Test

The airflow profile test uses quantitative techniques to evaluate the uniformity of air release across the surface of ceramic diffusers while operating rather than appraising uniformity by visual means. This is accomplished by testing the element at an air rate which is approximately equal to 2 SCFM/sq. ft., or at the recommended design rate, with anywhere from 2-8 inches of water over it.

The rate of air release from small areas is measured by displacing water from an inverted graduated cylinder and recording the rate of displacement of water with a stopwatch. The same sites monitored in the BRP test are tested in this procedure. By combining the cylinder area and the rate of air discharge, a flux rate, expressed as SCFM/sq. ft. or other convenient units, can be caluculated. By comparing the individual readings with one another, a quantitative measure or graphical representation of the profile can be generated.

The combination of DWP, BRP, and flow profile tests applied to new diffusers and at various stages in their operating history, provides a very useful diagnostic tool in evaluating the rate, the nature and the effect of fouling, be it organic or inorganic, on fine bubble porous diffusion elements. It is also effective in appraising the effectiveness of various cleaning procedures.

## DIFFUSER - FOULING

Porous diffuser plates were used in activated sludge processes as early as 1916 and they became the most popular method of aeration by the 1930's (8,9). Shortly after their installation, however, it became clear that clogging could be a problem. In a comprehensive review of fine bubble diffuser servicing, Roe, (10), outlined a number of causes of diffuser clogging. This list has been expanded over the years and is presented in Table I.

It was clear even in the early years of fine bubble diffusion that clogging was highly site specific and quite often difficult to forecast (9,11). Diffuser clogging was chronic at some plants after only a few months of operation, yet, at others, virtually no difficulties with clogging arose. Substantial effort was applied in the early years to reducing internal clogging by means of efficient air cleaning (12,13,14), a practice still performed today. External clogging was not as easy to analyze or control, however, and operations research concentrated primarily on treating the symptom rather than the cause.

### TABLE I. CAUSES OF DIFFUSER CLOGGING

Air Side

- Dust and Dirt from unfiltered air
- Oil from compressors or viscous air filters
- Rust and scale from air pipe corrosion
- Wastewater solids entering through diffuser or pipe leaks

Liquor Side

- Fibrous material attached to sharp edges
- Inorganic fines entering media at low or zero air pressure
- Organic solids entering media at low or zero air pressure
- Oils or greases in wastewater
- Precipitated deposits, including iron and carbonates
- Biological growths on diffuser media

In today's plants air side fouling has been all but eliminated except for a few cases where mixed liquor leaks into the system through a bad joint or broken pipe. In a few cases, however, we have observed the presence of fibers from air silencers on the internal surfaces of tubes and discs.

By and large, the majority of fouling that is observed today involves liquor side fouling. This fouling is conveniently divided into inorganic fouling, as a result of the precipitation substances including carbonates and iron, and biological fouling as a result of the development of biological growths or slimes on the fine pore media.

An excellent reference on inorganic clogging is a paper published by Morgan (15). He investigated and reported on clogging related to the presence of iron, calcium bicarbonate, fine sand in the mixed liquor and particulate matter in the air supply. Regarding iron he indicates information on iron clogging has confirmed that which was known from plant experience, that ferrous iron will result in external clogging. For plants adding iron salts for phosphorus removal this may be a problem and should be considered. Morgan indicates that clogging caused by calcium bicarbonate in the water system is not well understood, probably because it is not commonly encountered. Of these bicarbonates, only calcium is objectionable and is found in troublesome concentrations in few cities. In industrial applications, such as the pulp and paper industry, calcium fouling may be more likely to ocur.

Based on our experience and that of Houck and Boon (6), it appears that liquor side fouling related to the occurrence of biological growths on the media represents a much more significant problem than does the inorganic clogging. In the Houck and Boon survey (6), it is reported that the single greatest operational problem encountered in the survey was the formation of biological growths on the diffusers operating in zones of high volumetric loading and/or low dissolved oxygen. In only two cases, however, were domes observed to contain excessive amounts of biological slime, (Beddington, U.K., and Madison, WI). All other plants apparently exhibited signs of "coarse bubbling" which the authors attributed to the development of biological slimes. The authors further cited pilot plant work at the Water Research Centre at Stevenage, U.K., which verified the causes of diffuser sliming as the result of very high loading rates (12). No data was presented, however.

Boyle and Redmon (7), drawing from the literature and their experience, reported causes of biofouling, biofouling agents, the effects of biofouling and design for control of biofouling. Some of the suspected factors affecting biofouling are presented in Table 3.

TABLE III. SUSPECTED FACTORS AFFECTING BIOFOULING

- Low permeability of diffusers (16, 11)
- Low air flow rate per diffuser (12)
- High organic loading (6, 12)
- Low dissolved oxygen concentration (6)
- High temperatures (12, 14)
- Uneven air distribution among diffusers (10, 12, 16)
- Uneven air distribution within a diffuser
- Waste Characteristics

Boyle and Redmon (7), point out that the development of a biological growth on diffuser media could have several effects on diffuser performance. Two possible scenarios were given. One involved a gradual build up of biological growths where the bioslime leads to bubble coalescence and little noticeable change in DWP is observed until the slime envelopes the entire media surface. Under these conditions, the uniformity of the air released from the diffuser may be detrimentally affected and poor distribution may result. The OTE may decrease as bubble coalescence occurs and there could be visible changes in bubble patterns and the appearance of "coarse bubbling" at the surface. This type of diffuser cloggging may not be evident as routinely measured by air line pressure losses or demand for more air, (as a result of some loss in transfer efficiency). The second scenario involved biofouling simultaneous with internal clogging or external encrustation. In this case it was hypothesized that the larger pores would plug first resulting in the progressive utilization of the finer pores producing, perhaps, more uniform distribution and finer air bubbles. Increases in DWP could be small at this point. Eventually the biological mat may result in bubble entrapment and coalescence. The net effect of this progressive clogging might be an increase on OTE up to a point, (as air distribution improves and bubble size decreases), followed by a drop in OTE as coarser bubbles begin to form. Dynamic wet pressure would slowly rise to a point where only the finer pores are carrying air; after this, pressures would rapidly rise.

## DIFFUSER SURVEILLANCE

Considering that the nature and magnitude of fouling is highly site specific and difficult to forecast (9,11), it seems desirable to install a test header in an existing aeration basin to identify the rate of clogging and nature of the foulants. Using the procedures previously described, a rapid indication of the rate of clogging can be obtained. In addition, amenability to cleaning and definition of optimum cleaning methods can be explored.

Sanitaire, one of our clients, has done a significant amount of this testing. In some test installations the rate of pressure rise was very low and in some others, such as a potential industrial application, a 20-40 inch water gauge rise was observed over a period for four to six weeks. The latter case would be one where the use of fine pore media would not be a good selection.

Similar diffuser monitoring for an operating ceramic grid system can also be quite beneficial. This monitoring can be done on several diffusers in the operating system and/or in conjunction with a small test header. Under this condition one can characterize several ceramic elements for DWP vs. flow, BRP and airflow uniformity prior to installation and monitor DWP at a fixed air rate over a period of time to observe the rate of pressure increase. Figure 1 illustrates the pressure measurements needed to do this.

In conjuction with the observations of DWP with time, measurement of oxygen transfer efficiency with time can help identify the appropriate time for cleaning. One way to accomplish this is the measurement of local OTE and uniformity of air release by the offgas apprach. This technique will be described in Session 4.

Using data generated as above, the operator would have sensitive objective information upon which to determine the extent of fouling. Using a small test header in conjunction with the operating system permits visual observation of the foulants by removing the small header from the aeration basin. The degree from fouling and the most appropriate cleaning procedure can be ascertained from the test diffuser.

One apparent problem with many of the existing ceramic grid systems is that the existing monitoring methods of measuring line pressure are not sufficiently sensitive, and the operating personnel do not decide to clean the system until the system is substantially plugged. In this condition, it may be very difficult or impossible, to restore the system. Careful survelance in conjunction with an active maintenance program which attempts to keep the system in relatively new operating condition is considered a proceedure.

## DIFFUSER CLEANING

Depending upon the nature and extent of clogging as well as the diffuser materials and system make up, there are various cleaning methods available. A few methods that come to mind include sand blasting, brushing, steam cleaning, surface cleaning followed by acid soaking, refiring with or without prior acid soaking, ultrasonic cleaning, high pressure water sprays, and in-situ cleaning with acidic gases. Except for the last method, all the other methods require process interrruption.

The most popular techniques in present use seem to be cleaning the media surface by brushing, hosing, or the like, followed by acid soaking. Houck and Boon (6), report dome cleaning at the Beckton Plant in the United Kingdom, which includes surface cleaning, refiring and acid washing. The refiring is done on site in their refiring furnace. In the U.S., refiring has not been too prevalent, but some manufacturers are suggesting it in extreme cases. Whether this is a cost effective procedure in the U.S., where the ceramic elements must be shipped back to the manufacturer, remains to be seen.

A relatively recent innovative approach regarding cleaning, developed by Sanitaire (17), involves injection of gases which are aggressive to the fouling agents. In most cases, HCl gas is introduced into the supply air which results in an acidic reaction product when the gas contacts the water entrained in the element.

The philosophy of the gas cleaning approach is to carefully observe the dynamic wet pressure of the system and clean at such frequency to maintain the media in a like new condition. Cleaning at this frequency should prevent the system from ever becoming substantially clogged, thereby maintaining it's oxygen transfer efficiency and prolonging the life of the porous media. Several tests of this procedure have been tried on a small scale, and in the great majority has been very effective in reducing the system back pressure requirements and maintaining uniformity of air distribution.

When using this method, it is important that the materials of construction are resistant to the treating gas. Also of importance, is the uniformity of air release within the element and the distribution of air among the system, since the gas acts primarily at the active sites on the diffuser element. Thus, for a system with poor uniformity across the media and between diffusers, gas cleaning will preferentially clean those areas of highest flow and may not result in substantial improvement of system performance.

## FUTURE RESEARCH

The impact of long-term fine bubble diffuser operation in wastewater with respect to maintenance requirements and OTE is still poorly documented. It is clear that some biological growth will develop on diffuser surfaces but the impact of that growth on performance is unclear. Research should be directed under controlled experimental conditions toward measuring the progression of biological fouling employing sensitive BRP and air flow distribution techniques. These studies should include a variety of diffuser materials and geometry. The selection of diffuser systems is often predicated on the performance of clean diffusers in clean water. The long term changes in OTE during diffuser operation in wastewater should be more clearly delineated in order to provide the engineer with a clearer picture of what happens to OTE over the design life of the diffuser.

In addition, studies to determine the optimum pore size or specific permeability, that is the pore size which optimizes the rate of fouling and OTE, would be of real value.

As energy costs continue to escalate, engineers will seek more effective energy conserving systems for the transfer of oxygen to wastewater. Until a better understanding of the behavior of fine bubble diffusers under long term field conditions is realized, doubts will continue to be raised regarding the effectiveness of the fine bubble system. Futhermore, mis-application of these systems in certain situations may result in a greater expenditure of manpower and energy than necessary.

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# SELECTED EXPERIENCE WITH AERATORS USED IN THE TREATMENT OF PAPER INDUSTRY WASTEWATERS

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## INTRODUCTION

It was mentioned in a previous paper (1) that the selection of aerators for use in treating pulp and paper mill wastes is based on a series of considerations, only one of which is oxygen transfer. Many times it is of equal or greater importance that the machine selected perform reliably, require a minimum of maintenance and that it be a cost effective selection.

This paper presents the operational experience from eight out of over 400 plants which use aerators in treating paper industry wastewaters. Although only a few plants are presented, their problems and solutions are representative of those uncovered at other plants employing similar equipment during similiar periods of time. No attempt has been made to cover all of the types of aerators in use within the industry or all of the manufacturers which supply the industry with aeration equipment. Further, it is expected that improvements in machinery design, inspection procedures and operational controls provide today's purchasers with a more reliable machine for use in industry and municipal wastewater treatment systems than may have been the case in the 1970's.

## GENERAL EXPERIENCE

The paper industry has had considerable experience operating aeration equipment for the purpose of reducing the BOD of its wastewater. For example, it is estimated that over 200,000 horsepower are currently employed to aerate about 4000 MGD of wastewater in the U.S. paper industry.

In the paper industry, it is common practice to interview operations personnel at many sites before a choice of aeration equipment is made for a new site or an expanded site. These interviews provide the industry with up to date information on the performance of equipment as well as document problems in need of correction. In addition, the summaries of NCASI and other workshops often contain an organized review of operating and maintenance information which is useful in surveying the experience of others.

It should be mentioned that the paper industry employs mostly surface mechanical aerators in its wastewater plants. Therefore, the experience collected is largely directed toward this type of aerator.

## Gear Wear

Experience with operation of gear reducers on slow speed large diameter aerators is quite varied within the paper industry. For example, some units have operated with the original gear boxes for over twenty years, whereas, others require complete replacement because of wear on gears and bearings after only a short period of operation. This problem has been the most frequently reported difficulty encountered with aerators in the paper industry. Where gear systems fail, the cost of replacing the parts is often sufficient to consider purchasing a new aerator. The case histories presented later illustrate this practice.

## Fastening System Failures

Some fastening system problems reported in the paper industry have involved stationary mixers fastened to an air header resting near or on the bottom of the basin. The vibrations caused by turbulence in transporting air and water have been sufficient to cause fracture in some fastening systems and fatigue in others, resulting in losing the aerator.

Although not very prevalent in the paper industry, dual blade mechanical aerators have lost blades and experienced twisted shafts. The cause is usually related to improper fastening of the lower shaft resulting in uneven distribution of forces along the shaft.

Although only a fastening problem in the broadest sense, the action of ice on various structures which support the aerator can cause the development of uneven forces on the aerator shaft and in a few cases has caused the capsizing of floating aerators. In many locations where freezing is normal, pier corrals or cable supports have been provided to prevent floating units from tipping. Also, fixed mounted aerators, although not subject to turnover, can be shielded from accumulating large blocks of ice which, if loosened, could cause damage to an impeller.

## Seal Failures

High speed direct drive surface aerators are usually floated on the water surface. There has been only an occasional report of failure of the seal which protects the motor from contacting water on this type of device.

## Construction Material Failures

Paper industry effluents may contain clay and fly ash which are abrasive to certain materials of construction. Also, certain wastewaters may be corrosive to unprotected mild steel. Aerators may also be placed in natural basins where stumps and debris were never removed.

These wastewater and system characteristics have caused isolated problems, some of which are presented in the case histories which follow. In most cases, when surfaces are frequently inspected for wear and corrosion, the damage to concrete piers, steel pontoons and the like can usually be reduced by taking some preventative action.

## CASE HISTORIES

The paper industry is used to operating high speed mechanically complex equipment in processing pulp and manufacturing paper. Skilled personnel, well equipped shops and maintenance machines are available at most plants. Furthermore, the processes are operated continuously and personnel are familiar with making repairs during operating days in a timely manner. The maintenance of a mechanical aerator or a blower is not much different from other machinery, once its operating principles become familiar.

This portion of the paper assembles several case histories of aerator operational problems taken from contact with industry representatives. Most of the cases reviewed cover relatively long histories of experience with aerator operation. Although these situations don't cover every problem that might arise or represent the experience of the entire paper industry, it is hoped that they present the user's perspective which will be helpful to others operating these systems.

### Mill A

This bleached kraft mill designed its own aerator in the late 1950's which consisted of a top and bottom impeller located on a shaft which rested on the bottom of the basin. Each of four activated sludge basins eventually contained three 25 HP and two 20 HP units which were powered by a motor and gear reducer system. The mill is still operating these units, although after over 25 years of operation they are now being replaced with larger commercial aerators.

Mill personnel believed that these units required relatively high maintenance and are hoping to reduce that maintenance with the new units. The shop built units required frequent replacement of the foot bearing due to wear. Also, occasional shaft failures and gear reducer bearing failures were experienced. Although each gear reducer bearing performed with a different rate of wear, the average interval between replacements was about three years.

### Mill B

Mill B operates an activated sludge plant with about 800 stationary bubble mixers installed in a 2.5 acre basin which is

12 feet deep. During the first two years of operation, all of the original fiber glass mixers were worn out due to erosion from clay in the effluent. The underwater piping was also severely damaged when hold down straps cut through the PVC piping.

Replacement of the system during operation required division of the basin by a massive concrete wall and repiping of the basin so that each part could be operated independently. After the wall was in place, half of the basin was taken out of service, pumped down, and the aerators repaired. The PVC piping and fiber glass mixers were replaced with duplicate equipment manufactured from polyethylene. Problems with breakage of the mixer support legs were experienced and these legs have been strengthened or replaced.

Pluggage of the air distribution piping is a minor operating problem which increases with time. The underwater piping has been cleaned by divers periodically while the basin is in operation. Increasing air pressure due to plugging of the air distribution piping has caused a surging condition in the centrifugal air blowers. Surging can and did destroy the blowers. Prevention was easily obtained by installation of low amperage shutdown of the blower motor. Also, when down, the air piping is cleaned to reduce the back pressure.

## Mill C

A key element in Mill C's activated sludge process is its 17-foot deep aerated basin. Oxygen transfer is provided by 100 HP low (single) speed surface aerators. Adjustment in aerator performance is accomplished by adjusting a variable level overflow weir in the aerated basin. The three goals in aerator operation are:

(1) To maintain in suspension the required level of biological solids to accomplish BOD removal. Additional concern was expressed in minimizing any potential deposition in the aerated basin due to the carryover of solids from primary clarification. The activated sludge process at this location is treating a deink mill wastewater.

(2) To maintain sufficient dissolved oxygen in the mixed liquor of the aerated basin.

(3) To minimize the stress placed on the aerator gear box (reducer) by the varying hydraulic load resulting in a fluctuating submergence on the rotating aerator element.

It was pointed out that the above goals could be implemented by manipulating the adjustable overflow weir; however, the response time was too long, particularly in minimizing the "wear and tear" on the aerators. The alternate solution was to remove some of the aeration blades (scoops) to reduce amperage and gearbox stress. No reduction in process performance has been observed as a result of this modification. It should be noted, however, that residual dissolved oxygen in the aerated basin prior to the aerator modification was in the 4 mg/l to 5 mg/l range. This allowed the opportunity to sacrifice some oxygen transfer capability in an attempt to achieve goals 1 and 3 cited above.

## Mill D

Mill D is a largely intergrated kraft pulp and paper mill producing 550 tons per day of fine printing and specialty grades. The mill generates approximately 17 MGD of wastewater with an average untreated BOD of 40,000 pounds per day which is treated by a conventional actived sludge plant. The design flow and BOD to the plant is 21 MGD and 50,000 pounds per day, respectively.

Two rectangular basins each 400 ft. long by 120 ft. wide are operated in parallel. Aeration basin detention times are 9 to 10 hours at normal flows. Mixed liquor suspended solids are maintained at approximately 3500 mg/l with 50% of forward flow recycled back to the aeration basins.

Eight pedestal mounted mechanical aerators, four units per basin, provide aeration for the system. The electrical drives are two speed motors that have a nameplate output of 150 HP on high speed and 75 HP on low speed. The aerator gearbox provides a 38 to 1 speed reduction with a final output speed to the aerator impeller of 48 RPM on high speed and 32 RPM on low speed.

In general, it had been experienced that the gear reducers gave very reliable service, provided they were properly lubricated and cooled. Therefore, no spare gearboxes were ordered. Some problems were anticipated with the electrical drives and one standby unit was purchased. However, although no major problems developed with the electrical drives, five out of eight gearboxes failed since startup in May of 1976. Two units failed in January of 1978, and two more failed in May that same year. At some points, the replacement parts had not been received which resulted in having only 50% of design aeration capacity available.

An attempt was made to operate with only two aerators in each basin, but no matter how functional units were positioned, there were just too many dead zones in the basins and the effluent quality, particularly turbidity and suspended solids, became noticeably affected. Reducing the mixed liquor concentration had no detectable impact upon effluent quality.

During this period, mill maintenance crews became very skilled at jockeying aerators around so it was decided to divert all of the flow to one basin and switch all of the remaining aerators to that basin. This, in effect, reduced the detention time in the aeration basin to slightly more than four hours, but with the aerators on high speed there were no dead zones and effluent quality improved, but not to required levels. Operations continued in this mode for approximately one month until replacement parts for two units were received. With repair of these two units, the operational mode was changed to three aerators per basin and effluent quality returned to normal levels.

In August of 1979, just after replacement parts for the last two gearboxes had been received, a fifth unit failed. In all cases, the failures have been very similar; the teeth from the intermediate gears were stripped. Each time the gears were repaired, if the teeth on one gear were damaged, the gear to which it meshes, even though visibly undamaged, was replaced.

Shortly after the aerator gearboxes began to fail, personnel became concerned about the service life of the remaining units. For our particular system, several measures were taken:

(1) The aeration basin effluent weirs were lowered to their minimum attainable levels. This reduced the effective water level by 2.5 inches.

(2) The mixed liquor suspended solids were reduced by 1,000 mg/l and the recycled sludge rate was lowered to only 30% of forward flow.

(3) The aerator impeller shafts were shortened by 3 inches. This was the most effective measure taken, but also the most demanding in terms of mill maintenance crews.

(4) The aerator gearboxes were raised by about 1/2 inch by turning the adjustment nuts on the support bolts.

Prior to the development of the aerator gearbox failures, the aerators were set at an average submergence of five inches below the water surface, but after the above corrective measures were taken, the average aerator setting was one inch above the water surface. A recording wattmeter was placed on each aerator for a minimum of several days to insure that the aerator motor was drawing less than 150 HP.

After having spent many frustrating hours dealing with the problem of failing aerator gearboxes, mill personnel suggest the following guidelines for future installations:

(1) The aeration basins were equipped with DO probes connected to an automatic controller that would check DO levels every 20 minutes and change aerator speed from high to low or vice versa. This system should have a time delay relay to insure that when going from high speed to low speed the impeller shaft has stopped, or slowed down, to an RPM

less than slow speed before the motor actually switches to slow speed. It is felt that the lack of such a system is probably the most important single factor in the premature failure of our gearboxes.

(2) Each aerator should be started up with normal flow patterns and the submergence set by the use of a recording wattmeter. After each unit is set, it should be left connected to the wattmeter for at least 48 hours. All procedures should be clearly documented and saved for future reference.

(3) Each aerator should be equipped with its own ammeter. However, do not rely totally on ammeter readings.

(4) Consider the purchase of a complete set of internal gears and bearings to act as a spare.

(5) Be aware that the effective water level changes with flow. In this case, the difference between zero flow and average flow conditions was 2.5 inches.

(6) Finally, if any warranty agreement is reached that is even a little bit different than the standard warranty, make absolutely sure that these differences are very specific and in writing.

## Mill E

Mill E has recently installed a new aeration system in its activated sludge plant. After one year of operation, it was discovered that the lower shaft has sheared off at the flange below the surface impeller. These dual blade aerators contain a small mixing impeller at the base of the shaft. Mill personnel did not notice that the mixing impeller was no longer attached because plant performance was not affected. The shafts have been straightened or replaced and operators periodically check to see if they are still attached. The failure is believed due to improper fastening of the lower shaft during construction.

## Mill F

This mill has operated an aerated stabilization basin since 1964. The initial aerators were 60 HP slow speed units floated upon hollow steel pontoons. Maintenence of the pontoon surface to prevent corrosion was a continuous problem.

The gear reducers failed on several of the aerators beginning in 1967. Initially, these units were replaced with 20 HP high speed floating aerators and later with 50 HP high speed units. However, the first lot of 50 HP units were subject to major bearing problems which eventually led to the manufacturer installing an added shaft support to the bearing system which solved the problem.

When mill capacity increased in 1972, a group of 75 HP high speed units with stainless steel floats were installed. These units developed problems in keeping the impeller pinned to the shaft. Eventually, a double pin was used which solved this problem.

As of 1982, the mill currently is no longer using the original slow speed units. In their place most of the 20 HP and 50 HP units operating, and the load related to all recent increases in production is being satisfied with new 75 HP high speed units.

## Mill G

This mill utilizes an aerated stabilization basin which in 1971 utilized nine 125 HP floating slow speed units. The gearboxes started breaking down after 1.5 years operation. Problems with their repair by an outside contractor prompted mill personnel to make their own repairs. Two additional slow speed units were added in 1975.

In 1978, five 75 HP high speed aerators were installed which were anchored to the bottom by three or four cables. Since that time, these units have operated with few problems. One bearing burn-out occurred and occasionally an anchor cable breaks. Later, an additional group of 75 HP units replaced more of the original units as gear boxes wore out. The mill engineers believe this policy will continue as long as it is cost effective.

## Mill H

This bleached kraft mill installed its activated sludge system in 1960 which incorporated 12-60 HP dual speed fixed mounted aerators. Another six of the slow speed aerators were installed in 1966. The system was DO controlled with a 20-30 minute delay built into the switching system. Because of the wear caused to the aerator gearing system, the DO control was performed manually rather than automatically. A two hour cycle replaced the half-hour cycle.

The mill began replacing the old units in 1976 with high speed floating aerators because it was less expensive than replacing gear sets. Also, by this time, the motors on the old units were showing signs of severe corrosion.

In 1976, the mill also installed the first of 240 stationary bubble mixers. Initial experience with the bubble mixers produced fastener problems in that the bolts sheared at the plate fastener because the bolt size was 1/16" less than the hole in the plate. A tighter tolerence solved the problem. Also, the air headers required additional bracing to hold them in place. Since these early problems, only 14 aerators have been lost.

## SUMMARY

A general review of the types of problems requiring aerator replacement or overhaul has been presented with the aid of eight case histories which illustrate the class of problems encountered in aerating paper mill wastewater. The problems were classified as ones largely involving (a) gear wear or (b) the failure in systems which fasten the aerator to the tank, basin or air header. Additional problems involving using incompatable materials of construction and the failure of motor/shaft seals were occasionally reported.

## REFERENCES

(1) McKeown, J.J. "Notes for Workshop Toward an Oxygen Transfer Standard", Proc. Workshop Toward an Oxygen Transfer Standard, EPA-600/9-78-021, p. 228, (April 1979).

(2) "Operation and Maintenance of Aeration Equipment with Emphasis on Avoiding Mechanical Failures", Proc. of the 1979 NCASI Northeast Regional Meeting, 260 Madison Ave., N.Y., Special Report No. 80-03 (Feb. 1980).

## DISCLAIMER

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# PROPOSED STANDARD FOR MEASUREMENT OF OXYGEN TRANSFER IN CLEAN WATER

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## ABSTRACT

This standard was developed to measure the rate of oxygen transfer from diffused gas and mechanical oxygenation devices to relatively large volumes of water (up to 1 million gallons) typical of those found in the activated sludge wastewater treatment process. It is intended that this standard be referenced in engineering specifications for compliance testing.

This test method is based upon removal of dissolved oxygen (DO) from the water volume by sodium sulfite followed by reoxygenation to near the saturation level. The DO inventory of the water volume is monitored during the reaeration period by measuring DO concentrations at several points selected so that each point senses an equal tank volume. The method specifies a minimum number, distribution and range of DO measurements at each point.

The data obtained at each determination point are analyzed by a simplified mass transfer model to estimate the apparent volumetric mass transfer coefficient, $K_La$, and the saturation concentration, $C^*_\infty$. Nonlinear regression is employed to fit the model to the DO profile measured at each point during reoxygenation. In this way, estimates of $K_La$ and $C^*_\infty$ as are obtained at each point. These estimates are adjusted to standard conditions and the standardized oxygen transfer rate (mass of oxygen dissolved per unit time in clean water under standard conditions of temperature and pressure at an hypothetical concentration of zero DO) is obtained as the product of the average adjusted $K_La$ value, the average adjusted $C^*_\infty$ value and the tank volume. A procedure based on the clean water test results is prescribed for estimation of oxygen transfer rates in dirty water under process conditions.

Various components of power consumption are defined and methods for measurement of gas rate and power consumption by the oxygenation device are given. Energy efficiency of the oxygenation device is evaluated as the mass rate of oxygen transferred per unit power consumed.

## Measurement of Oxygen Transfer Rate in Clean Water

### 1. Scope

This method covers the measurement of the oxygen transfer rate OTR, as a mass of oxygen per unit time dissolved in a volume of water by an oxygen transfer system operating under given gas rate and power conditions. Methods for measurement at gas rate and power are also described. It is intended primarily to measure the rate of oxygen transfer from diffused gas and mechanical oxygenation devices to relatively large volumes of water (up to 1 million gallons) typical of those found in the activated sludge wastewater treatment process. Although the method is intended primarily for clean water, it is applicable to dirty water when the rate of oxygen demand is negligible.

The method is applicable to tests in full scale field facilities, in factory shop tanks and in small laboratory tanks and is valid for both completely mixed and incompletely mixed tanks. The primary result is expressed as the Standardized Oxygen Transfer Rate, (SOTR), a hypothetical mass of oxygen transferred per unit time at zero dissolved oxygen concentration, water temperature of 20°C and barometric pressure of 1.00 atm, under specified gas rate and power conditions. The results can be applied to estimate oxygen transfer rates in dirty water at process dissolved oxygen (DO) concentrations.

### 2. Summary of Method

The test method is based upon removal of dissolved oxygen (DO) from the water volume by sodium sulfite followed by reoxygenation to near the saturation level. The DO inventory of the water volume is monitored during the reaeration period by measuring DO concentrations at several determination points selected so that each point senses an equal tank volume. These DO concentrations may be either sensed *in situ* using membrane probes or measured by the Winkler method applied to pumped samples. The method specifies a minimum number, distribution and range of DO measurements at each determination point.

The data obtained at each determination point are then analyzed by a simplified mass transfer model to estimate the apparent volumetric mass transfer coefficient, $K_La$, and the saturation concentration, $C^*_\infty$. The basic model is described in Reference 1 and is given by

$$C = C_\infty^* - (C_\infty^* - C_0) \exp(-K_La\,t) \qquad 1$$

where

$C$ = DO concentration, m $L^{-3}$

$C_\infty^*$ = DO concentration attained as time approached infinity, m $L^{-3}$

$C_0$ = DO concentration at time zero, m $L^{-3}$

$K_La$ = Apparent volumetric mass transfer coefficient $t^{-1}$, defined so that $K_La = \dfrac{\text{rate of mass transfer per unit volume}}{C_\infty^* - C}$

Non-linear regression is employed to fit Equation 1 to the DO profile measured at each determination point during reoxygenation. In this way, estimates of $K_La$ and $C_\infty^*$ as are obtained at each determination point. These estimates are adjusted to standard conditions and the standardized oxygen transfer rate (mass of oxygen dissolved per unit time at an hypothetical concentration of zero DO) is obtained as the product of the average adjusted $K_La$ value, the average adjusted $C_\infty^*$ value and the tank volume. A procedure is also prescribed for estimation of oxygen transfer rates in dirty water under process conditions based on the measured value of SOTR in clean water.

## 3. Significance and Limitations

Oxygen transfer rate measurements are useful for comparing the performance and energy efficiency of oxygenation devices operating in clean water. However, performance of these devices in process water may significantly differ from the performance in clean water and the amount of difference will depend on the device and on the nature of the process water.

## 4. Definitions and Nomenclature

### 4.1 Power Terms

4.1.1 Delivered Power. Power delivered to the oxygen transfer device.

4.1.2 Delivered Gas Power. Power delivered to a subsurface gas injection device; calculated based upon adiabatic compression.

4.1.3 Delivered Shaft Power. Shaft power delivered to a turbine or other impeller; calculated based upon shaft torque and angular velocity measurements.

4.1.4 Pump Delivered Power. Water power delivered by a pump; calculated based on the volumetric flow rate, weight density and total dynamic head.

4.1.5 Total Delivered Power. Sum of delivered gas power plus, delivered shaft power plus pump delivered power.

4.1.6 Delivered Standard Air Power. Delivered gas power calculated based on adiabatic compression of air from 14.60 psia to 15.70 psia plus pressure loss through the diffuser psi plus hydrostatic pressure at diffuser submergence psi.

4.1.7 Delivered Standard Shaft Power. Shaft power required to turn on impeller in water at 20°C at an angular velocity equivalent to the actual angular velocity.

4.1.8 Pump Delivered Standard Power. Pump delivered power when the specific gravity of water is 1.00.

4.1.9 Total Delivered Standard Power. Sum of delivered standard air power plus delivered standard shaft power plus pump delivered standard power.

4.1.10 Wire Power. Electrical power drawn by a motor.

4.1.11 Wire Standard Power. Wire power drawn by an electrical motor of standard efficiency and transmitted by a system at standard efficiency to achieve a given level of total delivered standard power.

## 4.2 Mass Transfer Terms

4.2.1 Oxygen Transfer Rate (OTR). Mass of oxygen per unit time dissolved in a volume of water by an oxygen transfer system operating under given conditions of temperature, barometric pressure, power, gas rate and dissolved oxygen concentration.

4.2.2 Oxygen Transfer Rate at Zero DO ($OTR_o$). OTR when the DO concentration is equal to zero at all points in the water volume.

4.2.3 Oxygen Transfer Rate in Dirty Water ($OTR_f$). OTR for the oxygenation system operating in dirty water at a specified average DO concentration and temperature.

4.2.4 Standardized Oxygen Transfer Rate (SOTR). OTR in clean water when the DO concentration is zero at all points in the water volume, the water temperature is 20°C and the barometric pressure is 1.00 atm.

4.2.5 Aeration Efficiency (AE). OTR per unit total power input. Power input may be based either on delivered power or wire power.

4.2.6 Standardized Aeration Efficiency (SAE). SOTR per unit standard power input; may be based either on Total Delivered Standard Power or Wire Standard Power.

4.2.7 Oxygen Transfer Efficiency (OTE). Fraction of oxygen in an injected gas stream dissolved under given conditions of temperature, barometric pressure, gas rate, and DO concentration.

4.2.8 Oxygen Transfer Efficiency at Zero DO ($OTE_o$). OTE when the DO concentration is equal to zero at all points in the water volume.

4.2.9 Standardized Oxygen Transfer Efficiency (SOTE). $OTE_o$ when the water temperature is 20°C and the barometric pressure is 1.00 atm.

## 5. Apparatus and Methods

5.1 Tank. A tank or vessel is required to contain the water to which oxygen is transferred. The geometry and size will depend on the particular oxygenation system to be tested. Tank volumes may range from small laboratory vessels of a few liters to large tanks of over 1 million gallons.

5.2 Water. For determination of a Standardized Oxygen Transfer Rate, the water to which oxygen is transferred should be equivalent in quality to a potable public water supply. Further specifications of clean water are given in Section 6.5. This method may also be applied to measure an oxygen transfer rate in wastewater when the rate of oxygen demand is negligable.

5.3 Oxygenation Device. This method is applicable to a wide variety of oxygenation devices installed in the tank including but not limited to the following:

Surface Aerators: high speed, low speed and horizontal shaft rotors.
Subsurface Oxygenation Devices: diffused air, static tubes, submerged turbines and jet aerators.

The method can be applied to systems based on air oxygenation as well as to those based on oxygenation by high purity oxygen.

5.4 Sampling Devices. Submersible pumps and tubing are necessary when DO concentrations are to be measured on pumped samples in accordance with Section 6.10.2.

## 5.5 Dissolved Oxygen Measurement

5.5.1 Wet chemical measurement of DO on pumped samples shall be in accordance with the Azide Modification of the Winkler Method described in Section 421B of the 15th Edition of Standard Methods (5) except when a blank correction is required as described in Section 6.11.3 in which case the corresponding procedure given in the 11th Edition of Standard Methods (6) should be used.

5.5.2 Membrane Electrode Measurement of DO either on pumped samples or *in situ* shall be in agreement with Section 421F of Standard Methods (5).

5.6 Temperature Measurement. Water temperature measurement shall be in accordance with Section 212 of Standard Methods (5).

## 5.7 Deoxygenation Chemicals

5.7.1 Sodium Sulfite. Technical Grade sodium sulfite ($Na_2SO_3$) shall be used for deoxygenation in accordance with Section 6.8. It is preferable that the sodium sulfite be free of cobalt. However, chemical containing a known concentration of cobalt may be employed provided that this cobalt is considered as part of the total cobalt addition discussed in Section 6.9.1.

5.7.2 Cobalt Catalyst. Either reagent grade or technical grade cobalt chloride, $CoCl_2$, or cobalt sulfate, $CoSO_4$, shall be used to catalyze the deoxygenation reaction in accordance with Section 6.8.

5.8 Electronic Computer. A digital computer or microcomputer capable of handling one of either the Fortran or Basic languages is required for the recommended non linear regression method of parameter estimation described in Section 7.2.1.

5.9 Gas Flow Measurement Apparatus. For oxygenation systems based on sub-surface gas injection, an apparatus capable of measuring the gas flow with an accuracy of ± 5% in accordance with Section 6.3 is required.

5.10 Power Measurement Apparatus. The apparatus required for power measurements will depend on the specific oxygenation device but, in general, apparatus suitable for measurement of Total Delivered Power and Total Wire Power in accordance with Section 6.4 is required.

## 6. Procedure

6.1 Advance Preparation and Responsibilities. When this method is to be applied, the engineer-owner-manufacturer (EOM) representatives shall agree in advance on the specific system to be tested and the test conditions. Agreement must be reached on

test location: field installation or shop tank
tank size and geometry
aerator placement
aerator power and gas rates

Where field testing is to be conducted, the engineer-owner representative should provide the manufacturer with detailed drawings and specifications of the tank or tank section in which the test will be conducted. Information on the water supply source and available water chemistry data should be provided. Water samples should be made available to the manufacturer for laboratory experiments regarding the chemical additions that will be made.

Once the installation of aeration equipment is completed, provision should be made for EOM representatives to inspect the installation to verify placement and testing conditions. Systems employing diffused air aeration should be tested to eliminate leaks. Provisions for power and air flow measurement should be verified and modifications made as needed. It may be necessary to install equipment such as meters for power measurement, supplemental air piping, orifice plates and manometers.

Upon completion of the installation of the aeration equipment, the test tank should be cleaned prior to filling for testing. Once the tank is filled with the test water, chemical and biological contamination should be avoided. It may be necessary to dewater and refill the test tank during the testing and adequate pumping and discharge arrangements should be made.

6.2 Test Tank Geometry and Aerator Placement. It is difficult to describe a required geometry or placement for testing conducted in tanks other than the full-scale field facility. Appropriate configurations for shop tests should as closely simulate the field conditions as possible. Water depths should be similar if not identical and interferences due to wall effects and any extraneous piping or other materials in the tank should be minimized. The density of aerator placement, air flow per unit volume or area and power input per unit volume are examples of parameters that can be used to assist in making comparative evaluations.

Testing of tank sections is also useful in certain situations (e.g., long, narrow diffused aeration tanks) where there is little water circulation between adjacent sections. In this approach, a tank is divided into sections and each section is tested individually. When this testing is performed, partitions shall be installed between adjacent sections to prevent interchange of oxygen by advection and diffusion. This technique can provide information on spatial variation of $K_La$ and SOTR in tanks designed for tapered aeration.

Consideration should be given to utilization of shop testing or testing of tank sections when full-scale facilities are very large (e.g., in excess of 1 million gallons). Other criteria to be considered in making this judgement are:

Ease of distribution of deoxygenation chemicals; distribution may be difficult in certain tanks.

Sampling requirements; very large tanks may be difficult to sample adequately.

Bulk flow and mixing patterns; shop or section testing should not be done when the mixing pattern in the shop or section tank would not be representative of the full scale unit.

6.3 Gas Flow Rate Measurement. For aeration systems based on subsurface gas injection, the gas flow rate shall be accurately and precisely determined. Recommended references for procedural assistance are Spink, L. K., Principles and Practices of Flow Meter Engineering (8), Cusick, C. F., Flow Meter Engineering Handbook (9), and A.S.M.E. Power Test Codes (10).

Full-scale plant gas flow measurement systems should be used with caution when conducting field tests in a portion of the plant. The precision and accuracy of the measurement device may not be adequate for test flow rates. Estimating gas flow rates to part of the plant by volumetric or area served ratios could be used as a check but not for the primary air flow test information.

Primary gas flow measurements shall be made using calibrated orifice plate meters, venturi tube meters, pitot traversing methods, e.g., the Annubar method, or other meters with an accuracy of ± 5 percent. It is desirable to provide a back-up or supplemental measurement system as a check. The instantaneous gas flow rate should not vary by more than ± 5 percent during the test.

The gas flow measurement system shall be installed to avoid any potential pulsation effect from positive displacement blowers. Pulsation can cause significant error in head meter readings and is evidenced by meter oscillations, inability to zero the meter under a no flow condition, or meter readings being affected by connecting tubing length. According to Spink, it can be mitigated by additional volume and head loss between the blower and meter.

The following data shall be obtained to relate the measured gas flow to standard conditions and to calculate the oxygen mass rate:

Measurements in flowing gas feed line:

Pressure, $P_f$
Temperature, $T_f$
Gas composition for gases other than air

Ambient Measurements:

Pressure, $P_a$
Temperature, $T_a$
Relative humidity, $R_h$

Ambient pressure and relative humidity may be obtained from Weather Service data.

Standard conditions for air flow are defined as 20°C, 1.00 atm, and 36 percent relative humidity. For gases other than air, the standard condition is defined as 20°C, 1.00 atm and zero humidity. For air, the volumetric flow rate at standard conditions is related to the actual volumetric flow rate in the feed line by,

$$Q_s = 36.2 \left[ \frac{P_f \left( 1 - \frac{R_h P_{va}}{P_a} \right)}{T_f} \right] Q_t \qquad 2$$

where: $Q_s$ = air flow rate at standard conditions, ft$^3$/min

$Q_t$ = actual air flow rate in the feed line, ft$^3$/min

$P_f$ = total pressure in the air feed line, psia

$T_f$ = air temperature in the feed line, $^oR$

$R_h$ = relative humidity of ambient air, dimensionless fraction at $T_a$

$P_{va}$ = vapor pressure of water at ambient temperature, $T_a$, psia

$P_a$ = ambient pressure, psia

Taking air to be 23 percent oxygen by weight, the mass flow rate of oxygen is given by,

$$W_{O2} = 1.034\, Q_s \qquad 3$$

where: $W_{O2}$ = mass flow of oxygen in air stream, lb/hr

$Q_s$ = volumetric air flow rate, scfm

6.4 Power Measurements. Power expended during an unsteady-state oxygen transfer test should be determined. Various power terms are defined in Section 4 and it is important that the specific components of power to be measured are clearly specified.

The power components to be measured depend on the specific oxygenation device but, in general, the total delivered power (power delivered to the oxygen transfer device) and total wire power (electrical power drawn by motors driving the device) should be measured.

Measurement of power and related quantities should be in accordance with the following:

Shaft power measurement - ASME Power Test Code PTC 19.7 (10)

Electrical power measurement - ASME Power Test Code PTC 19.6 (10)

Pressure measurement - ASME Power Test Code PTC 19.2 (10)

Rotary speed measurement - ASME Power Test Code PTC 19.3 (10)

### 6.4.1 Delivered Power

6.4.1.1 Total Delivered Power is given as the sum of all the delivered power terms.

$$\begin{matrix}\text{Total}\\ \text{Delivered}\\ \text{Power}\end{matrix} = \begin{matrix}\text{Delivered}\\ \text{Gas}\\ \text{Power}\end{matrix} + \begin{matrix}\text{Delivered}\\ \text{Shaft}\\ \text{Power}\end{matrix} + \begin{matrix}\text{Pump}\\ \text{Delivered}\\ \text{Power}\end{matrix} \qquad 4$$

Specific oxygenation devices may have only one or two delivered power terms, e.g., a surface aerator has only shaft power whereas a submerged turbine has gas plus shaft power.

6.4.1.2 Delivered Gas Power shall be calculated based on adiabatic compression by

$$\begin{matrix}\text{Delivered Gas}\\ \text{Power}\end{matrix} = \frac{WRT_1}{K}\left[\left(\frac{P_2}{P_1}\right)^{k} -1\right] \qquad 5$$

where the symbols and corresponding dimensions are given by

$w$ = weight flow rate of gas, $f \cdot t^{-1}$

$R$ = gas constant, $L\,T^{-1}$

$T_1$ = absolute temperature before compression, $T$

$P_1$ = absolute pressure before compression, $f\,L^{-2}$

$P_2$ = absolute pressure after compression, $f\,L^{-2}$

$k$ = ratio of specific heats for gas, $C_p/C_v$, dimensionless

$K$ = $\frac{k-1}{k}$, dimensionless

6.4.1.3 Delivered Shaft Power should be based on direct measurement of torque and angular velocity delivered to the oxygenation device and calculated from

$$\begin{matrix}\text{Delivered}\\ \text{Shaft}\\ \text{Horsepower}\end{matrix} = 2\pi\, T_q \,\frac{N_r}{3.3 \times 10^4} \qquad 6$$

where the symbols and corresponding units are given by

$T_q$ = torque. ft·lb, measured using either a cradled dynamometer (cradled motor, generator, prony brake) or surface strain dynamometer (surface strain, angular twist)

$N_r$ = rotational speed, revolutions per minute

When accurate conversion efficiencies are known, Delivered Shaft Power may be calculated by multiplying the motor wire or motor shaft horsepower by the appropriate efficiency.

6.4.1.4 Pump Delivered Power shall be calculated as

$$\text{Pump Delivered Horsepower} = \frac{Q_w\,(\text{T.D.H.})(\text{S.G.})}{3960} \qquad 7$$

where the symbols and corresponding units are given by

$Q_w$ = liquid flow rate, gal/min

T.D.H. = total pump head, ft

S.G. = specific gravity of water at test temperature

6.4.2 Delivered Standard Power. For purposes of comparison, it is convenient to base delivered power on standard conditions.

6.4.2.1 Delivered Standard Air Power is the gas power consumed by the oxygen transfer device operating on water at 68°F (20°C) when fed standard air at a mass rate of oxygen equivalent to the actual air feed rate with a blower intake pressure loss of 0.10 psi and a pressure loss in the blower discharge line of 1.00 psi. This shall be calculated by

$$\text{Delivered Standard Air Horsepower} = 0.227\, Q_a \left[ \left( \frac{P_{2s}}{P_{1s}} \right)^{0.283} - 1 \right] \qquad 8$$

where:

$Q_a$ = air rate expressed as scfm

and $P_{2s}$ and $P_{1s}$ refer to standard discharge and inlet pressures such that

$$P_{2s} = \text{standard atmospheric pressure (14.70 psia)} + \text{hydrostatic pressure at the diffuser, } P_h + \text{diffuser pressure loss } P_d + \text{standard pipe pressure loss (1.0 psia)} = 15.70 \text{ psia} + P_h + P_d$$

$$P_{1s} = \begin{array}{l}\text{standard}\\ \text{atmos-}\\ \text{pheric}\\ \text{pressure}\\ \text{(14.70 psia)}\end{array} - \begin{array}{l}\text{standard}\\ \text{suction}\\ \text{pressure}\\ \text{loss}\\ \text{(0.1 psia)}\end{array}$$

$$= 14.60 \text{ psia}$$

6.4.2.2 Delivered Standard Shaft Power is the shaft power required to turn the turbine in water at 20°C at an angular velocity equivalent to the actual angular velocity. This should be estimated from Equation 6 based upon the variation of shaft torque with water temperature at constant angular velocity.

6.4.2.3 Pump Delivered Standard Power is the pump delivered power given by Equation 7 when the specific gravity is 1.00.

6.4.2.4 Total Delivered Standard Power is given by

$$\begin{array}{l}\text{Total}\\ \text{Delivered}\\ \text{Standard}\\ \text{Horsepower}\end{array} = \begin{array}{l}\text{Delivered}\\ \text{Standard}\\ \text{Air H.P.}\end{array} + \begin{array}{l}\text{Delivered}\\ \text{Standard}\\ \text{Shaft H.P.}\end{array} + \begin{array}{l}\text{Pump}\\ \text{Delivered}\\ \text{Standard}\\ \text{H.P.}\end{array} \qquad 9$$

6.4.3 Wire Power. Wire power is related to delivered power by

$$\text{Wire Power} = \frac{\text{Delivered Power}}{e_a\, e_b\, e_c \, \ldots} \qquad 10$$

where:

$e_a$, $e_b$, $e_c$ are the efficiencies of the various power conversion devices (motors, gear drives, reducers, blowers) used to transform electricity to delivered power.

Wire power should be measured using accurate recording polyphase watt meters capable of monitoring 10 cycle per second peaks. An ammeter can also be used if the voltage and power factor are measured. An expression for calculating the 3-phase power from current, voltage and power factor measurements is

$$\begin{array}{l}\text{Wire}\\ \text{Horsepower}\end{array} = 2.319 \times 10^{-3}\ E\ I\ F \qquad 11$$

where:

E = voltage, volts

I = current, amperes

F = power factor

6.4.4 Wire Standard Power. Wire Standard Power is useful when efficiency comparisons are to be made on the basis of wire power. In certain test situations, blower or motor efficiency may not be typical of the efficiency attained in an actual system. Consequently, actual wire power might be representative of a typical installation. Wire Standard Power shall be calculated by applying the recommended standard efficiencies given in Table 1 to the appropriate Delivered Standard Power Term.

### 6.5 Water Quality

6.5.1 General and Total Dissolved Solids. The water supplied for the initial test shall be equivalent in quality to a potable public water supply. Subsequent tests may be conducted in the same water and, because of the addition of deoxygenation chemicals, the Total Dissolved Solids concentration (TDS) will increase. Repetitive testing may be conducted in the same water provided that the TDS does not exceed 1500 mg/l. When the water supply contains TDS greater than 500 mg/l, the 1500 mg/l limit may be raised to 2000 mg/l with the approval of EOM representatives.

6.5.2 Temperature. Water temperature should be between $10^oC$ and $30^oC$ and as close to $20^oC$ as possible. Testing outside this temperature range may be necessary in some field situations and can be done with the approval of EOM representatives. Low temperatures slow the deoxygenation reaction and this may introduce some error. It is recommended that a standard $\theta$ value of 1.024 be employed to adjust for temperature. Appreciable error can be introduced when the actual $\theta$ value differs from this and the temperature difference is more than 5 degrees Celsius. Water temperature shall not change by more than 2 degrees Celsius during a single unsteady state test.

6.5.3 Water Quality Analyses. Initial Analyses: Prior to beginning the testing program, a representative sample of the water to be used in the test tank shall be analyzed for: TDS, alkalinity, sulfate, iron, manganese, residual chlorine, pH, total organic carbon or chemical oxygen demand, cobalt, surfactant (MBAS), and temperature. EOM representatives should review these data to assess possible effects of water quality on the test results. Based on these data, EOM representatives should establish: 1) upper limits on the allowable TDS or sulfate concentrations; 2) the concentration of cobalt to be employed; 3) possible modification to the dissolved oxygen measurement procedure e.g. use of a blank in Winkler analyses; and 4) possible modifications of the test water quality.

Normally, no modifications of test water quality are made for clean water testing. However, it may be that, due to a particular set of field conditions, a surface water may have to be treated. Demonstrated oxygen transfer rate effects due to water chemistry differences between the field conditions and prior test data used to predict the field performance may be considered in agreeing upon any water quality modifications.

Table 1. Recommended Equipment Efficiencies to be Used for Standard Power Determinations$^{(3)}$

| Item | Efficiency Range (decimal) | Recommended Standard Efficiency (decimal) |
|---|---|---|
| Compressor | 0.50 - 0.80 | 0.70 |
| Gear box | 0.94 - 0.96 | 0.95 |
| Coupling or drive | 0.90 - 1.00 | 0.95 |
| Motor | 0.90 - 0.95 | 0.92 |

*Efficiencies quoted are for full load conditions.

Table 2. Water Quality Measurements During Testing

| Measurement | Frequency of measurement |
|---|---|
| Temperature* | every test - beginning and end |
| Total Dissolved Solids | every test or calculated by mass balance |
| Filterable Cobalt | every test, end |
| pH* | every test |
| DO Demand*+ | every test plus early in test when DO < 50 percent saturation |

*measured at test site during testing.
+optional, to be performed when unreacted sulfite is suspected.

Water Quality Analyses During Testing: When repetitive tests are conducted in the same water, the water should be analyzed for the various constituents and properties at the frequencies indicated in Table 2. The DO demand measurement is made by filling a BOD bottle with sample and measuring the oxygen depletion at 1 and 5 minutes. This measurement is intended to indicate residual unreacted sulfite. A significant DO demand by unreacted sulfite seriously reduces the apparent transfer rate and will invalidate the results.

The cobalt concentration normally should not change during testing but slight decreases have been observed. If the concentration of cobalt falls below 0.1 mg/l, additional cobalt should be added to assure complete oxidation of the sulfite.

When detergent-added testing is conducted, analyses for anionic surfactant and surface tension should be made at intervals during each run. It has been reported that the concentration of surfactant decreases during a test. Allowing for a 20 minute delay after addition of detergent and prior to deoxygenation may assist in stabilizing the surfactant concentration.

6.6 System Stability. The aeration system should be operated to achieve steady state hydraulic conditions prior to starting the oxygen transfer evaluation. The hydraulic mixing regime should be established in the test tank for each test condition prior to deoxygenation. A steady power draw can indicate a steady state hydraulic condition for mechanical aerators. Some mechanical and diffused air systems require 30 to 40 minutes to achieve a steady state hydraulic regime.

For diffused air systems, water shall be displaced from the aeration system prior to beginning the test. Steady manometer readings for orifice air flow measurement and consistent air flow rate measurements for other flow measurement devices are indicative of this displacement. Lines with values for purging water from the aeration system may be added for testing purposes.

For tests of fixed platform surface aerators and rotors, the water surface elevation shall be held constant so that the power draw is constant during the test. For other systems, the volume of water under aeration shall not vary by more than ± 2 percent during any one test.

6.7 Initial Run. Data anomalies have frequently been reported during the initial run conducted after filling the test tank. Because of this, this run is frequently not used as part of the oxygen transfer evaluation. Instead, it is often used to stabilize water chemistry and to overcome possible problems in test procedures including sulfite dispersion and sampling techniques. It also can provide an opportunity to check for possible analytical interferences, to assure proper probe calibrations and to verify adequate cobalt residual based on deoxygenation and reaeration patterns. The initial run may be used as part of the oxygen transfer evaluation with the approval of EOM representatives.

6.8 Deoxygenation Chemicals. Technical grade, sodium sulfite ($Na_2SO_3$) shall be used for deoxygenation. The sulfite shall be essentially cobalt free and contain no impurities that would alter the oxygen transfer rate analysis. Sodium sulfite should be added in solution or slurry form. This is accomplished by dissolving the sulfite in a separate mixing tank prior to its addition to the test tank. Saturated solutions contain 2.23 lb/gal at 20°C and 3.00 lb/gal at 30°C.

Technical grade sulfite from a given batch should be evaluated for impurities and potential interference in oxygen uptake evaluations. This should be done by chemical analysis for cobalt, iron and manganese and by comparative laboratory oxygen transfer tests in which the results using the supplied technical grade sulfite for deoxygenation are compared against those obtained using analytical grade sulfite or nitrogen stripping deoxygenation methods.

The sulfite deoxygenation reaction is catalyzed by cobalt. The cobalt source utilized should be cobalt chloride, $CoCl_2$, or cobalt sulfate, $CoSO_4$, reagent or technical grade. The cobalt should be dissolved prior to its addition to the test tank. Solubilities of $CoCl_2$ and $CoSO_4$ are approximately 50 g and 36 g per 100 ml of water respectively at 20°C and increase with temperature. Care should be taken to ensure that complete dissolution of the cobalt salt is obtained, especially when it is added in the sulfate form.

6.9 Addition of Deoxygenation Chemicals

6.9.1 Cobalt Addition. A solution of cobalt salt shall be added to the test tank to achieve a soluble cobalt concentration between 0.10 mg/l and 0.50 mg/l in the test water. Cobalt concentrations greater than 0.50 mg/l, but less than 2.0 mg/l are permissible when laboratory aeration experiments demonstrate that greater concentrations are necessary to catalyze the deoxygenation reaction in a particular test water or when the water temperature is less than 10°C. If a cobalt concentration greater than 0.5 mg/l is employed, analytical precautions should be taken to eliminate potential interferences in the Winkler titration. Soluble cobalt measurements should be made at the conclusion of each test and these results may indicate that supplemental additions of cobalt are required.

The cobalt solution shall be added prior to the beginning of oxygen transfer testing with the aeration system operating. The solution shall be uniformly distributed into the test tank. Pumps and distribution systems may be required for large tanks. The cobalt solution shall be dispersed throughout the tank by operating the aeration system for approximately 30 minutes. The cobalt catalyst should normally be added once for each test water.

6.9.2 Sulfite Addition. The theoretical sodium sulfite requirement for deoxygenation is 7.88 mg/l per 1.0 mg/l DO concentration. Sulfite additions are made in excess of stochiometric amounts. The amount of excess is dependent on the oxygen transfer rate of the aeration system and the size of the test tank. The amount of excess varies from 20 to 150 percent and increases for high transfer rate systems.

Sufficient sulfite solution shall be added to depress the DO level below 0.50 mg/l at all points in the test water. It should be noted that consistent repetitive testing results have been observed where the DO concentration has reached zero at all sample points and remained at zero at least two minutes. Results from the initial test run can be used to help establish the proper quantity to be added.

Sodium sulfite shall be dissolved or slurried in mixing tanks outside the test tank and distributed uniformly and rapidly (in less than 3 to 4 min) into the test tank. The use of pumps and flexible piping to distribute the solution across the tank surface is recommended. Testing conditions may prevent achieving a true sulfite solution prior to the addition. Addition in a slurry form is preferred over the direct addition in a crystal form. Extreme care should be exercised to assure adequate dispersion and dissolution in the test tank.

6.10 Determination of Dissolved Oxygen at Various Points in the Tank During the Unsteady State Test. Dissolved oxygen concentration shall be determined at various points in the tank and at various times during the unsteady state test. This determination shall be carried out by one or both of the following procedures:

Samples pumped to BOD bottles followed by measurement of dissolved oxygen by either the Winkler method or membrane probe.

In situ measurement of dissolved oxygen in the tank by membrane probes.

6.10.1 Location of Dissolved Oxygen Determination Points. The number and location of dissolved oxygen determination points will be dictated by the size of the test tank, aerator placement and mixing pattern in the tank. The following criteria shall be used in establishing the number and location of points.

A minimum of four determination points shall be used. One should be at a shallow depth; one should be at a deep location and one should be a mid-depth. The points should be at least 2 feet from the walls, floor and surface and no closer to the surface than 10 percent of the minimum tank dimension.

The determination points shall be located so that each senses an equal portion of the tank volume and shall be distributed vertically and horizontally to best represent the tank contents.

> More than four determination points should be used when the tank is large (e.g. > 500,000 gal), when significant dissolved oxygen gradients exist (e.g. more than 1.0 mg/l DO difference between points), or when a point source oxygenation device is installed in a large tank.

For large tanks with point sources of oxygen and non-uniform dissolved oxygen concentrations, as many as 8 determination points may be required.

Occasionally, a DO probe will malfunction during a test and, for this reason, it is advisable to use backup probes near selected determination points so that the test data may be valid even if one probe malfunctions. Backup probes should be clearly identified as such, and the determination points to which each backup probe applies should be specified.

6.10.2 Collection of Pumped Samples. When pumped samples are to be obtained for dissolved oxygen analysis, submersible pumps (discharging through tubing to BOD bottles) should be installed at the determination points. The pump inlet should be designed and located to avoid air bubble extrainment. The pumping rate and tubing lengths should be adjusted so that the transport times between the pump and bottle are equal for all determination points and not greater than 15 seconds, and the 300 ml BOD bottles fill in 6 to 10 seconds. Sudden pressure drops in the sample line will degas the water and thus should be avoided.

With this system, the DO measurement may be made by a DO probe mounted in the sample line, by a DO probe with a stirrer placed into the sample bottle or by Winkler titration of the DO in the sample bottle. The in-line probe system must be observed continuously to prevent line clogging or damage to the membrane. If individual bottle samples are used, they must be carefully stored to prevent temperature change and degassing. They should be analyzed for DO as soon as possible.

If the pumped sample-individual bottle technique is used, it is recommended that at least one recording DO probe system also be installed. This system would be used to help set sampling times, indicate when zero DO was reached during deoxygenation and assist in determining the end of run by noting a stable maximum DO concentration.

6.10.3 Times of Dissolved Oxygen Determination. A minimum of 10 to 15 dissolved oxygen values shall be measured at prescribed times at each determination point during the unsteady state test. Measurement of up to 20 values is desirable but more than 20 values will not usually result in a significant improvement in precision. Dissolved oxygen determination shall be made at times so that two-thirds (63 to 70 percent) of the values are evenly distributed over the period of time between the initial data point which is above zero and $2/K_La$ (0 to 86 percent saturation) and one-third (30 to 37 percent) of the values evenly distributed over the period of time from $2/K_La$ to $4/K_La$ (86 percent to 98 percent saturation). In cases of rapid transfer, the minimum interval between measurements should be 0.4 minutes for pumped samples. No minimum interval is specified for *in situ* probes.

It is convenient to plan a test based on either 12, 15, 18 or 21 measurements. Table 3 shows the determination intervals estimated to give the required distribution of data points for these cases. To use this table for calculation of intervals, one must have an a priori estimate of either $K_La$ or OTR. Furthermore, the table assumes that sampling will begin at time zero which is assumed to occur at zero dissolved oxygen concentration. If zero DO is not attained or if early truncation is to be practiced, the determination interval for zero to 86 percent saturation should be decreased by roughly 25 percent so that the required number of points are obtained in this region.

6.10.4 Run Duration and DO Saturation. DO data should be obtained over as wide a range as possible. Truncation of data at DO levels less than 20 percent of $C^*_\infty$ is allowable to avoid lingering effects of the deoxygenation technique. In no case shall values of DO greater than 30 percent of $C^*_\infty$ be truncated.

All test runs shall be continued for a period of time approximately equal to 4 divided by the anticipated value of $K_La$. This is equivalent to continuing the run until the DO concentration is 98 percent of the saturation concentration, $C^*_\infty$, which is normally approached at a time equal to $6/K_La$. For purposes of comparison with the value of $C^*_\infty$ estimated by the model, it is useful that at least one run be continued a time at $6/K_La$ for each testing condition (temperature, aerator operation and geometry) to obtain a test saturation DO concentration.

Measured values and tabulated values of dissolved oxygen surface saturation concentrations shall be used for comparative information only and shall not be used as model parameters for calculation of oxygen transfer rates.

### 6.11 Dissolved Oxygen Measurements

6.11.1 Measurement by In Situ and Sample Line DO Probes. The in situ DO probes shall be fast response probes with 1.0-mil membranes and should be equipped with agitators. If agitators are not employed, care shall be taken to ensure that there is sufficient water velocity past the probe. The probes shall be calibrated using the Winkler procedure with test tank water and checked for linearity against Winkler procedure titrated samples. The calibration and linearity shall be established prior to the initial run and shall be verified every 3 to 4 hours during subsequent runs. It is recommended that a calibration and linearity check be performed on at least one probe during each run. Probe calibration and linearity check can be conveniently accomplished by comparing probe readings with Winkler measurements on discrete samples taken at the probe locations. Each probe should be equipped with a strip chart or digital recorder which will permit reading of the dissolved oxygen concentration at 0.5 minute intervals with a precision of ± 0.05 mg/l. However, in the absence of recording devices, it is permissible to manually record probe data at the appropriate determination intervals.

Table 3. Estimated Dissolved Oxygen Determination Intervals

| | Time from 0 to $2/K_La$ (0 to 86% Saturation) | | Time from $2/K_La$ to $4/K_La$ (86% to 98% Saturation) | |
|---|---|---|---|---|
| Total Number of Data Points | Number of Points | Determination Interval, min. | Number of Points | Determination Interval, min |
| 12 | 8 | $17.1/K_La$ | 4 | $30/K_La$ |
| 15 | 10 | $13.3/K_La$ | 5 | $24/K_La$ |
| 18 | 12 | $10.9/K_La$ | 6 | $19.8/K_La$ |
| 21 | 14 | $9.2/K_La$ | 7 | $17.1/K_La$ |

$K_La$ = volumetric transfer coefficient, $hr^{-1}$ and may be approximated by

$$K_La = \frac{OTR_o}{10W}$$

where: $OTR_o$ = expected oxygen transfer rate at zero dissolved oxygen, lb/hr

W = weight of water, $10^6$ lb

If the probes are placed in sample lines to measure DO, they need not be equipped with agitators. The velocity of flow past the probe shall be established and maintained to provide an accurate response from the probe. Calibration, linearity check, and recording requirements for sample line probes are equivalent to those for in situ probes.

Dissolved oxygen probes require considerable care and attention to provide continuous reliable data. Back-up probes and meters should be available for replacement as needed.

6.11.2 Measurement by Probes in BOD Bottles. Probes used to measure the DO in 300-ml sample bottles shall be equipped with a stirrer. Probe manufacturer recommendations shall be followed for calibration and linearity checks.

6.11.3 Measurement by the Winkler Procedure. The modified Winkler DO analytical procedure of the 15th Edition of Standard Methods (Section 421 B) should be followed for DO titrations. Procedures using excess iodide to prevent the loss of iodine vapor at high DO concentrations should be considered if unexplained lower readings in the higher DO ranges are experienced.

Chemical interferences in Winkler DO titrations have been observed with some testing, especially when higher cobalt levels are used (in the 2.0 mg/l range). The analytical procedure presented in the 11th edition of Standard Methods (1960) should be used to correct for these interferences when cobalt concentrations greater than 1 mg/l are present. A sample of the test water should be taken at the end of the test run and titrated directly without the addition of manganous sulfate solution, but with the addition of the alkali-iodide-azide reagent. This value, called a "blank", should be subtracted from every DO measurement in the test tank to correct for chemical interferences.

6.12 Testing with Detergent Addition. The unsteady state clean water oxygen transfer test is occasionally performed with clean water containing added detergent to roughly simulate the performance of the oxygen transfer device in a dirty water and to mask the effect of trace contaminants in tap water. The transfer rates obtained in the presence of detergents are not necessarily equivalent to those in municipal wastewater or mixed liquor. However, the transfer rates in the presence of detergents are generally closer to those obtained in actual process water than are the usual clean water transfer rates.

A strong solution of detergent, linear alkylate sulfonate (LAS) or househould detergents, shall be prepared using hot water (80-90°C). The detergent solution shall be added in an amount to achieve an average concentration of approximately 5 mg/l (measured as methylene-blue active substance-MBAS) during a test run. An initial concentration of 7 mg/l of detergent in the test water has been used. The detergent solution shall be added prior to the reaeration period and shall be uniformly dispersed throughout the tank contents.

The detergent concentration (MBAS) in the test water shall be analyzed at the beginning and end of the run and at several intermediate points to determine an average concentration for the run. Static surface tension measurements at these times are also recommended.

Because cobalt may be complexed by detergents, the cobalt concentration should be checked to verify that adequate catalyst is present. It is suggested that a cobalt addition equivalent to 0.10 mg/l in the test water be made for each run. Provision should be made to dispose of any foam that develops during the test. Some difficulty with detecting end points with the Winkler DO titration procedure has been noted. When detergents are present it is recommended that probes be used to measure the DO concentration. The remaining test procedures are as noted in earlier sections.

## 7. Data Analysis

### 7.1 Preparation of Data for Analysis.

The purposes of data preparation are:

to obtain discrete numerical values from continuous probe readings,

to plot and examine the data to determine if low-end truncation is necessary,

to truncate low dissolved oxygen values, and

to assure that the data values to be analyzed meet the timing criteria outlined in Section 6.10.3.

7.1.1 Continuous Probe Data. When DO probes equipped with continuous strip chart recorders are used, a continuous record of dissolved oxygen versus time will be available at each determination point. In this case, approximately 21 discrete data values shall be selected from the recorder output for each determination point at the approximate time intervals given by Table 3. If the probes are equipped with digital print out, the discrete values closest to the interval divisions shall be selected. The initial data value shall be at the beginning of record and shall be greater than zero, but less than 0.5 mg/l dissolved oxygen. The discrete data values shall be read directly from the recorder graph or digital output and shall not be subject to any smoothing or averaging procedure. If low-end truncation is anticipated or if the record does not begin at zero DO, the determination interval for 0 to 86 percent saturation (0 to $2/K_La$) should be decreased from the value of $9.2/K_La$ indicated in Table 3 to about $7/K_La$ so that 14 or more points are selected in the zero to 86 percent range. Approximately seven discrete data values shall be selected in the range of 86 to 98 percent saturation ($2/K_La$ to $4/K_La$) at intervals of approximately $17.1/K_La$.

Use of Table 3 requires an approximate value of $K_La$ and one way of obtaining this is based on the expected value of $OTR_o$ as indicated in the Table. However, an approximate value of $K_La$ can more easily be established based on inspection of the DO versus time plot by noting the approximate value of the saturation concentration, $C^*_\infty$, approached at infinite time. An approximate value of $K_La$ is then given as the reciprocal of the time interval between dissolved oxygen concentrations of zero and 63 percent of $C^*_\infty$. (For this method to be applied, the data may have to be extrapolated to zero DO).

The 21 or more data values selected at each determination point should be plotted as DO versus time, and the low DO data should be examined for lingering effects of the deoxygenation chemicals. These effects would be manifested by the rate of increase in DO (dC/dt) being constant or increasing with DO (C) up to an inflection point. If this pattern is suggested, the DO data below 20% of the approximate $C^*_\infty$ value, low DO data may be truncated up to a DO concentration of 1.5 times the DO concentration at the inflection point. However, in no case shall DO values greater than 30 percent of $C^*_\infty$ be truncated. DO values in the 86 to 98 percent saturation range shall not be truncated except as required in the application of the Alternate Best Fit Log Deficit parameter estimation method (Section 7.2.2).

Following possible low-end truncation, the data plotted for each determination point should be checked again for compliance with the timing criteria outlined in Section 6.10.3. In the case of data obtained from recording probes, 63 percent to 70 percent of the DO values (at least 13) shall be evenly distributed over the period of time between the initial data point and $2/K_La$ (where the DO is 86 percent of $C^*_\infty$) and 30 to 37 percent of the DO values (at least 6) shall be evenly distributed over the period of time from $2/K_La$ to $4/K_La$ (86 percent to 98 percent of $C^*_\infty$).

Occasionally, a DO probe will malfunction and data from a malfunctioning probe shall note be used to calculate the SOTR. Frequently, a malfunction will be obvious but, in other cases, a malfunction may be only suspected. The following criteria should be employed to judge whether data from a suspected probe may be deleted.

- Results of a probe calibration or check conducted immediately before, during, or immediately after the test.
- Comparison of the $K_La_{20}$ value for the suspected test and point with the average of those obtained for the same point in at least three replicate tests (including the suspected test) as described in Section 8.2.2. Variation by more than 15% of the average suggests probe malfunction.

Deletion of data from a malfunctioning probe may invalidate the test in accordance with Section 8.2.2.

7.1.2 Discrete Pumped Sample Data. When discrete pumped samples are analyzed either by the Winkler method or by a membrane probe, the data consist of dissolved oxygen concentration values at the determination intervals used for each sampling location. In this case, it is crucial that the determination intervals or sampling times have been well planned in accordance with Table 3.

The data shall be plotted as DO versus time and the low DO data should be examined for lingering effects of the deoxygenation chemicals as described in Section 7.1.1 above. These effects are normally more difficult to detect in discrete sample data. Based on this examination, data values up to 20 percent or 30 percent of $C^*_\infty$ may be truncated as outlined in Section 7.1.1. DO values greater than 30 percent of $C^*_\infty$ shall never be truncated. DO values in the 86 to 98 percent saturation range shall not be truncated except as required in the application of the Alternate Best Fit Log Deficit Method (Section 7.2.2).

Following possible low-end truncation, the data plotted for each sampling point should be checked for compliance with the timing criteria outlined in Section 6.10.3. In the case of data obtained from discrete samples, 63 to 70 percent of the DO values (at least 7) shall be evenly distributed over the period of time between the initial data point and $2/K_La$ (where the DO is 86 percent of $C^*_\infty$) and 30 to 37 percent of the DO values (at least 3) shall be evenly distributed over the period of time from $2/K_La$ to $4/K_La$ (86 percent to 98 percent of $C^*_\infty$). The discrete data values to be analyzed shall not be subjected to any smoothing or averaging procedure.

7.2 Parameter Estimation. The purpose of the parameter estimation procedure is to determine the best estimates of the three model parameters, $K_La$, $C^*_\infty$, and $C_0$ so that the model given by Equation 1 best describes the variation of DO with time at each determination point location in the tank. Parameter estimates shall be performed for the data obtained at each determination point location by one of two methods, the recommended non-linear regression method or the alternate best fit log deficit method. The parameter estimates given by each method will differ slightly because of the top end truncation and logarithmic transformation required in the alternate method. However, the values of the transfer rate ($OTR_0$) calculated based on the parameter estimates will generally agree within a few percent. Other parameter estimation methods such as the ordinary log deficit or direct methods can be useful in determining approximate values for $K_La$ and $C^*_\infty$. However, the parameter estimates based on these other methods shall not be used for calculation of the transfer rates ($OTR_0$ and SOTR).

7.2.1 Recommended Non Linear Regression Method. This method is based on non linear regression of the model (Equation 1) through the DO versus time data as prepared for analysis in Section 7.1. The best estimates of the parameters $K_La$, $C^*_\infty$ and $C_0$ are selected as the values which drive the model equation through the prepared DO concentration versus time data points with a minimum residual sum of squares. That is, the parameter estimates are selected so that the sum at the squares of the residuals is minimized. Here, a residual refers to the difference in concentration between a measured DO value at a given time and the DO value predicted by the model at the same time.

Application of this method requires either a digital computer or microcomputer capable of handling one of either the Fortran or Basic Languages. A user-oriented Fortran program which will give the least square estimates, and standard deviations of $K_La$, $C^*_\infty$ and $C_0$ along with an output of the data, fitted values and residuals is given in Appendix A. A similar user-oriented Basic program written for the Apple II Microcomputer is given in Appendix B. To use either program, the prepared data for a particular determination (sampling) point are entered along with initial approximations of the three parameters. The computer then adjusts these initial approximations to minimize the residual sum of squares and reports the final parameter estimates*.

The values of DO predicted by the model based on the final parameter estimates should be plotted along with the measured values and examined. A visual examination of such a plot will indicate any gross mistakes in the non linear regression.

7.2.2 Alternate Best Fit Log Deficit Method. This method is acceptable when lack of a computing facility prevents application of the recommended method. The chief advantage of this approach is that it can be applied by using an ordinary scientific calculator (a calculator such as the Texas Instruments Model 55 with linear regression capability is recommended) instead of a computer. However, this method can be time-consuming and tedious.

The method is based on linear regression of a logarithmic form of the model equation using a logarithmic function of the DO data. The logarithmic form of the model equation is

$$\ln(C_\infty^* - C) = \ln(C_\infty^* - C_0') + K_L a\, t_0 - K_L a\, t \qquad 12$$

where:

$t_0$ = time when $C = C_0'$

This equation shall be fit to the prepared data for each determination (sampling) point by performing a linear regression of $\ln(C_\infty^* - C)$ versus time. The value of $C_\infty^*$ shall be estimated by trial and error. The best estimate shall be the value of $C_\infty^*$ giving the minimum residual sum of squares. The best estimate of $K_L a$ shall be determined as the slope of the linear regression based on the best estimate of $C_\infty^*$. Because of the lack of precision in very small values of $(C_\infty^* - C)$ and impossibility of taking the log of negative values of $(C_\infty^* - C)$, it is necessary to truncate (delete) values of C within 0.2 to 0.3 mg/l of $C_\infty^*$ when this parameter estimation method is applied. To apply this method, the following steps shall be followed.

1. The prepared data plotted according to Section 7.1.1 or 7.1.2 shall be examined and an initial trial value of $C_\infty^*$ determined by inspection of the plot.
2. Data points within 0.2 to 0.3 mg/l of the trial value of $C_\infty^*$ shall be truncated. However, the highest DO value included in the analyses shall be at least 92 percent of the trial $C_\infty^*$. In the case of continuous probe data, at least 3 values shall be at DO values greater than 86 percent of the trial $C_\infty^*$. For discrete sample data, at least 2 values shall be at DO values greater than 86 percent of the trial $C_\infty^*$.

*Note: Because of the way in which $C_0$ is defined in the model, it is possible to have negative values for the best estimate of $C_0$.

3. The quantity, $\ln(C^*_\infty\text{-}C)$ shall be computed for each remaining DO value and a linear least squares regression of this quantity versus time shall be made. (See any standard statistics text or suitable electronic calculator instructions). The regression coefficient is the slope of the plot of the $\ln(C^*_\infty\text{-}C)$ vs. time and is equal to $-K_La$. The value of $K_La$ corresponding to the trial value of $C^*_\infty$ shall be recorded. Also the value of the y intercept (point at which the regression line crosses the $\ln(C^*_\infty\text{-}C)$ axis at t = 0) shall be recorded. The predicted values of $\ln(C^*_\infty\text{-}C)$ shall be calculated for the time of each observation by

   $$\ln(C^*_\infty - C)\Big|_{\text{predicted}} = \text{y Intercept} - K_La\,t \qquad 13$$

   The residuals shall be calculated for each data point by

   $$\text{Residual}_n = \ln(C^*_\infty - C)\Big|_{\text{Measured}} - \ln(C^*_\infty - C)\Big|_{\text{predicted}} \qquad 14$$

   Finally, the residual sum of squares corresponding to the trial value of $C^*_\infty$ shall be calculated for all N data points as

   $$\begin{array}{l}\text{Residual}\\ \text{Sum of Squares}\end{array} = \sum_{1}^{N} (\text{Residual}_n)^2 \qquad 15$$

4. Steps 2 and 3 shall be repeated for at least three more trial values of $C^*_\infty$ in an effort to located the value of $C^*_\infty$ which minimizes the Residual Sum of Squares. The minimum residual sum of squares shall be established by plotting the Residual Sum of Squares as the ordinate versus the trial value of $C^*_\infty$. A number of trial $C^*_\infty$ values shall be selected to establish the value of $C^*_\infty$ which minimizes the residual sum of squares. Values of the residual sum of squares cannot be compared when based on different numbers of data points. Consequently, the truncation criterion (0.2 to 0.3 mg/l) shall be chosen so that the calculations for successive trial values of $C^*_\infty$ are based on the same number of data points.

5. The best estimate of $C^*_\infty$ shall be the value which minimizes the residual sum of squares. The corresponding value of $K_La$ shall be determined from the linear regression based on that value of $C^*_\infty$.

## 8. Interpretation and Reporting of Results

8.1 Standarized Oxygen Transfer Rate (SOTR). By convention, the oxygen transfer capacity of an oxygenation system is usually expressed as the rate of oxygen transfer predicted by the model at zero dissolved oxygen under standard conditions of temperature and pressure, usually 1.00 atmosphere and 20°C. This is termed the Standardized Oxygen Transfer Rate (SOTR). It should be noted that the SOTR is a hypothetical value based on zero dissolved oxygen in the oxygenation zone which is not usually desirable in real oxygenation systems operating in process water. The SOTR value shall be determined by correcting the values of $K_La$ and $C^*_\infty$ estimated according to Section 7.2.1 or 7.2.2 for each determination point to standard conditions by

$$K_La_{20} = K_La\,\theta^{(20-T)} \qquad 16$$

$$C^*_{\infty 20} = C^*_\infty\left(\frac{1}{\tau\,\Omega}\right) \qquad 17$$

where:

$K_La$ = determination point value of $K_La$ estimated according to Section 7.2.1 or Section 7.2.2.

$K_La_{20}$ = determination point value of $K_La$ corrected to 20°C

$\theta$ = empirical temperature correction factor, defined by Equation 16; shall be taken equal to 1.024 unless proven to have a different value for the aeration system and tank tested (see Section 9.3).

$C^*_\infty$ = determination point value of $C^*_\infty$ estimated according to Section 7.2.1 or Section 7.2.2.

$C^*_{\infty 20}$ = determination point value of $C^*_\infty$ corrected to 20°C and a standard pressure of 1.00 atm.

$\tau$ = temperature correction factor = $\dfrac{C^*_{st}}{C^*_{s20}}$

$C^*_{st}$ = tabular value of dissolved oxygen surface saturation concentration, mg/l, at the test temperature and a standard total pressure of 1.00 atm, (5)

$C^*_{s20}$ = tabular value of dissolved oxygen surface saturation concentration, mg/l, at 20°C and a standard total pressure of 1.00 atm, shall be taken as 9.07 mg/L (5)

$\Omega$ = pressure correction factor = $\dfrac{P_b + \gamma_{wt} d_e - P_{v20}}{P_s + \gamma_{ws} d_e - P_{v20}}$

$P_b$ = barometric pressure during test, $f/L^2$.

$P_{v20}$ = saturated vapor pressure of water at 20°C.

$P_s$ = standard barometric pressure of 1.00 atm, $f/L^2$.

$\gamma_{wt}$ = weight density of water at the test water temperature, $f/L^3$.

$\gamma_{ws}$ = weight density of water at 20.0°C, $f/L^3$.

$P_{vt}$ = saturated vapor pressure of water at the test temperature, $f/L^2$.

$d_e$ = effective saturation depth at infinite time, defined by

$$d_e = \frac{1}{\gamma_W}\left[\frac{C^*_\infty}{C^*_{sT}}(P_s - P_{vT}) - P_b + P_{vT}\right]$$

The average values of $K_La_{20}$ and $C^*_{20}$ shall be calculated by averaging the values at each of the n determination points by

$$\text{Average } K_L a = \overline{K_L a_{20}} = \frac{1}{n}\sum_{1}^{n} K_L a_{20} \qquad 18$$

$$\text{Average } C^*_{\infty 20} = \overline{C^*_{\infty 20}} = \frac{1}{n}\sum_{1}^{n} C^*_{\infty 20} \qquad 19$$

The Standard Oxygen Transfer Rate (SOTR) shall be computed by

$$\text{SOTR} = \overline{K_L a_{20}}\ \overline{C^*_{\infty 20}}\ V \qquad 20$$

where: V = volume of water in the test tank

The individual and average values of $K_La_{20}$, $C^*_{\infty 20}$, $d_e$, and the actual test temperature and tank volume shall be reported along with the SOTR. For subsurface gas injection systems, the value of SOTE should also be reported (See Section 8.4). If possible, the standard deviations of the $K_La$, $C^*_\infty$, parameter estimates should also be reported.

8.2 Spatial Uniformity and Reproducibility of $K_La$ and $C^*_\infty$ Values. In some cases, replicate tests are conducted sequentially under the same conditions of temperature and pressure and the replicate $K_La$ and $C^*_\infty$ values can be compared directly without temperature and pressure adjustments. However, when temperature and pressure vary between replicates, comparisons should be made using the $K_La_{20}$ and $C^*_{\infty 20}$ values.

8.2.1 Spatial Uniformity of $K_La_{20}$ and $C^*_\infty 20$ Values Related to Mixing and Data Adequacy. Uniformity of the point $K_La_{20}$ and $C^*_\infty 20$ values between the various determination points is a necessary condition for theoretical complete mixing with respect to dissolved oxygen concentration and oxygen transfer. Non-uniformity of these parameters indicates that theoretical complete mixing is not achieved. Although non-uniformity does not necessarily invaldiate the test, it does suggest non-uniform aeration intensity or short circuiting of the flow with some zones of the tank achieving higher oxygen transfer rates than others. Therefore, when non-uniformity is evident, care should be taken to ensure that a sufficient number of determination (sampling) points are adequately located to correctly sense the changes in dissolved oxygen inventory of the tank.

The spatial variation of average point $K_La_{20}$ values determined from the results of at least three replicate tests can be used to judge the adequacy of determination point numbers and locations. When a minimal number of determination points (e.g. four) are employed, variation of these average point values should be limited so that three-fourths of the values are within ± 10% of the mean value for the tank. When spatial variations greater than this are observed consideration should be given to using a greater number of determination points (e.g. six to eight) or to testing by tank sections.

8.2.2 Reproducibility of Point $K_La_{20}$ Values as a Criterion of Validity. When a series of at least three replicate tests are conducted, the point $K_La_{20}$ values determined in each replicate shall not vary by more than ± 15% from the mean value for that point. Replicate point $K_La_{20}$ values which exhibit greater variation shall be considered as invalid and shall not be used for calculation of the measured SOTR. Deletion of data from a determination point shall invalidate the replicate test unless suitable data from a backup probe are available for that test.

8.3 Standardized Aeration Efficiency. Frequently, the Standarized Aeration Efficiency (SAE) or rate of oxygen transfer per unit power input is of interest and shall be computed from:

SAE = SOTR/Power Input 21

This parameter is normally expressed in units of pounds per horsepower hour or kilograms per kilowatt hour.

8.4 Oxygen Transfer Efficiency (OTE). Oxygen transfer efficiency (OTE) refers to the fraction of oxygen in an injected gas stream dissolved under given conditions. Frequently, the Standardized Oxygen Transfer Efficiency (SOTE) which refers to the OTE at a given gas rate. Water temperature of 20°C, and barometric pressure of 1.00 atm is calculated. For a given flow rate of air, this is given by

$$\mathrm{SOTE} = \frac{\mathrm{SOTR\ lb/hr}}{1.034\ Q_s} \qquad 22$$

where: $Q_s$ = volumetric air flow rate, scfm

8.5 Performance Evaluation Criteria. Required oxygen transfer performance should be stated as a required standardized oxygen transfer rate (SOTR) for specified conditions of tank volume, placement geometry, power and gas flow-rate. System performance under the specified conditions shall be evaluated by a minimum of three replicate tests. The measured SOTR for performance evaluation shall be determined as the average SOTR based on all valid replicate tests under the specified conditions, and this average shall equal or exceed the required value. Furthermore, the individual SOTR values determined from at least two-thirds of the valid replicate tests shall exceed the required value.

## 9. Application of Clean Water Test Results to Estimate Oxygen Transfer Rates in Dirty Water at Process DO Levels.

The SOTR value determined by clean water non-steady state oxygen transfer tests may be applied to estimate the oxygen transfer rate, $OTR_f$, for the same oxygenation system operating in the same tank under the same conditions, but in dirty water at an average process level DO concentration and temperature. The estimated value of $OTR_f$ shall be calculated by

$$OTR_f = \frac{\alpha(\mathrm{SOTR})\theta^{(T_f-20)}}{\overline{C^*_{\infty 20}}} \cdot (\tau_f\,\beta\,\Omega_f\,\overline{C^*_{\infty 20}} - \bar{C}) \qquad 23$$

where: $OTR_f$ = oxygen transfer rate estimated for the system operating in dirty water at an average DO concentration $\bar{C}$ and temperature $T_f$

$T_f$ = temperature at the dirty water, °C

$$\alpha = \frac{\text{average dirty water } K_L a}{\text{average clean water } K_L a} = \frac{\overline{K_L a_{f20}}}{\overline{K_L a_{20}}}$$

$\tau_f$ = temperature correction, factor = $\frac{C^*_{sf}}{C^*_{s20}}$

$C^*_{sf}$ = tabular value of dissolved oxygen surface saturation concentration, mg/l, in clean water at the temperature of the dirty water. See Standard Methods (5) Table 421:I.

$\beta$ = $\frac{\text{dirty water } \overline{C^*_\infty}}{\text{clean water } \overline{C^*_\infty}}$

$\Omega_f$ = pressure correction factor $\frac{P_{bf} + \gamma_w \overline{d}_e - P_{v20}}{P_s + \gamma_w \overline{d}_e - P_{v20}}$

$C^*_{s20}$ = tabular value of DO surface saturation concentration at 20°C, shall be taken as 9.07 mg/l.

$\overline{d}_e$ = average effective depth, calculated as the average of the $d_e$ values measured at the clean water test determination points = $\frac{1}{n} \sum_{1}^{n} d_e$

$P_{bf}$ = barometric pressure for the dirty water application

$\overline{C}$ = dissolved oxygen concentration averaged over the entire dirty water volume

Other symbols are as defined in Section 8.2.

Although the application of clean water SOTR values to estimate transfer rates in dirty water is conceptually straightforward, the estimate of $OTR_f$ is subject to considerable uncertainty because of the uncertainty contained in the $\alpha$ value. This uncertainty is magnified when the dirty water application is based on tank geometry and temperature which differ from those of the clean water test.

Table 4 is a guide to the application of Equation 23 and indicates the source of information for the parameters needed to estimate $OTR_f$. Values of $\overline{C^*_{\infty 20}}$, $\overline{d}_e$ and SOTR must be known from the clean water test. The average DO value $\overline{C}$ must be determined from the dirty water conditions and should be evaluated as the process level DO concentration averaged over the entire aeration volume. It should not be taken as the DO concentration in the influent to a point source aerator. The temperature correction factor, $\tau_f$, and pressure correction factor, $\Omega_f$, should be calculated based on the definitions following Equation 23.

### Table 4. Guide to the Application of Equation 23

| Parameter | Source of Information |
|---|---|
| $\overline{C}^*_{\infty 20}$ | clean water test results |
| $\overline{d}_e$ | clean water test results |
| SOTR | clean water test results |
| $\overline{C}$ | Given by the dirty water conditions |
| $T_f$ | Given by the dirty water conditions |
| $\tau_f$ | Calculated based on tabulated DO surface saturation values |
| $\Omega_f$ | Calculated based on $\overline{d}_e$ and barometric pressure |
| $\alpha$ | estimated based on experience and on measured $K_La$ values |
| $\beta$ | Calculated based on total dissolved solids measurements |
| $\theta$ | Taken as 1.024 unless experimentally proven to differ |

9.1 Alpha Factor, $\alpha$. The alpha factor represents the ratio of the $K_La$ in process wastewater to the $K_La$ in clean water, and this ratio can range from approximately 0.4 to 1.2. It is influenced by a great number of process conditions including surfactants, tubulence, power input per unit volume, geometry, scale, bubble size, degree of treatment, and other wastewater characteristics. Ideally, the alpha factor would be measured by conducting full scale oxygen transfer tests with clean water and wastewater, but this is normally impractical. Several studies have described small scale (less than 50 gallons) oxygen transfer tests for measurement of the alpha factor and a state of the art method based on these studies has been recommended by Stenstrom (4). If alpha factor measurements in small scale vessels are necessary, the method given by Stenstrom should be followed. However, alpha factor measurement in small scale vessels is, at best, only an educated guess. In selection of an alpha factor for use in Equation 23, it should be borne in mind that, for a given wastewater stream, the alpha factor is normally not constant and a possible range of alpha values should be considered.

9.2 Beta Factor, $\beta$. The beta factor is defined as the ratio of the average saturation concentration, $\overline{C}^*_\infty$, in wastewater to the corresponding value in clean water. This ratio can vary from approximately 0.8 to 1.0 and is generally close to 1.0 for municipal wastewaters. Because it cannot be measured by a membrane probe and because many wastewaters contain substances which interfere with the Winkler method, it is difficult to measure accurately. For this reason, the value of beta for use in Equation 23 shall be calculated as the ratio of DO surface saturation concentration in the waste to the DO surface saturation concentration in clean water. The corresponding surface saturation concentrations shall be interpolated from Table 5 based on the total dissolved solids content of the wastewater and clean water. Table 5 is adapted from Table 421:I of Standard Methods (5) by converting the chloride to TDS for seawater.

9.3 Theta Factor, $\theta$. The theta factor is employed to correct $K_La$ for changes in temperature according to Equation 16. Values of theta reported in the literature have ranged from 1.008 to 1.047 and are influenced by geometry, turbulence level and type of aeration device. There is little consensus regarding the accurate prediction of theta values, and for this reason, clean water testing for the determination of SOTR values should be at temperatures close to 20°C.

The value of theta shall be taken equal to 1.024 in Equations 16 and 23. Use of a theta factor differing from 1.024 shall be permissible only when experimental data for the particular aeration system indicates conclusively that the value of theta is significantly different from 1.024.

Table 5. Dissolved Oxygen Surface Saturation Concentrations, mg/l at Various Temperatures and Approximate Total Dissolved Solids Concentrations. Adapted from Standard Methods (5).

| | Total Dissolved Solids mg/l | | | |
|---|---|---|---|---|
| Temp. °C | 0 | 9100 | 18,200 | 27,200 |
| 10 | 11.27 | 10.65 | 10.05 | 9.49 |
| 15 | 10.07 | 9.53 | 9.01 | 8.53 |
| 20 | 9.07 | 8.60 | 8.16 | 7.73 |
| 25 | 8.24 | 7.83 | 7.44 | 7.06 |
| 30 | 7.54 | 7.17 | 6.83 | 6.49 |

## ACKNOWLEDGEMENTS

This proposed standard is based on nearly four years of work by the ASCE Subcommittee on Oxygen Transfer Standards. A significant portion of the Committee's work was supported by grants from the U.S. Environmental Protection Agency.

Preparation of a standard general enough to be applied to all clean water unsteady state tests and specific enough to incorporate all essential procedures was difficult. Users of this standard must give particular attention to use of the mandatory "shall" and advisory "should" terms. For particular applications of this standard, it may be advantageous for the user to elevate certain advisory steps to the mandatory level. It is intended that this standard be referenced in engineering specification for compliance testing. When this is the case, the engineer should consider the costs of requiring extensive compliance testing in relation to the initial cost of the oxygen transfer system and present worth of future operating costs.

Several engineers, owners and manufacturers have developed sound oxygen transfer testing procedures and the procedures incorporated into this standard were judged to be the most accurate and generally applicable of those available. The substance of this standard is based on recommendations made in the report, "Development of Standard Procedures for Evaluating Oxygen Transfer Devices" by the A.S.C.E. Oxygen Transfer Standards Subcommittee, W. C. Boyle, Chairman. The user is referred to this document which contains references 1 to 4 and to reference 7 for background information.

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## DISCLAIMER

The work summerized herein was previously reviewed and approved for publication by the U.S. Environmental Protection Agency (EPA-600/2-83-002).

## Appendix A
## Fortran Program for Non Linear Regression

The Fortran computer program which follows in this Appendix is a non linear estimation of the parameters in the exponential form of the oxygen transfer model. The program uses the Taylor series linearization method and is in two parts. For greatest user ease, the computational and output portions are written in subroutine form (subroutine KLANL). The user supplied main program serves as an interface between the system constraints of the user's computer and the regression computational algorithm, in the subroutine.

The MAIN program must perform the following tasks. A sample MAIN program is shown in the Fortran listing.

1. Read in the data to be fitted; concentration (C) versus time (T).
2. Read in initial parameter estimates (CS, CO, XKLA).
3. Provide a descriptive name for the data set (INAME).
4. Provide the logical device number for output (NOUT).

Subroutine KLANL performs all the computations for the nonlinear estimation and controls the output of the results. The notation in the subroutine is consistent with that in this Appendix. The estimation and output proceeds in the following steps.

Step 1. Write titles, name of data set and headings for the iterations.
Step 2. Initialize internal variables and compute the fitted values and residual sum of squares based on the initial parameter estimates.
Step 3. Set up the normal equations using current parameter estimates.
Step 4. Solve normal equations for the corrections to the parameter estimates.
Step 5. Update parameter estimates and calculate new fitted values and residual sum of squares.
Step 6. Test for convergence. The convergence criteria are:
  a. Relative change in parameters less than 0.00001.
  b. Relative change in sum of squares less than 0.000001.
  c. The algorithm will also exit from the iteration loop if more than 10 iterations are required for convergence. A diagnostic message is printed.
Step 7. Calculate the estimated standard deviations of the model parameters.
Step 8. Write out a summary of the data, fitted values (F) and residuals (R).

An example estimation problem follows the Fortran listing.

```
C
C        MAIN PROGRAM FOR NONLINEAR ESTIMATION OF OXYGEN
C                   TRANSFER PARAMETERS
C
      DIMENSION C(100),T(100),F(100),R(100)
      WRITE(5,1)
    1 FORMAT('  NAME OF INPUT DATA FILE:  ' $)
      READ(5,2) INAME
    2 FORMAT(A5)
      OPEN(UNIT=20,MODE='ASCII',ACCESS='SEQIN',FILE=INAME)
      READ(20,10) CS,CO,XKLA
   10 FORMAT(3F)
      N=1
   15 READ(20,10,END=19) T(N),C(N)
      N = N+1
      GO TO 15
   19 CONTINUE
      NOB = N-1
      NOUT = 5
      CALL KLANL(C,T,F,R,NOB,CS,CO,XKLA,INAME,NOUT)
      CALL EXIT
      END
C
C     SUBROUTINE KLANL PERFORMS NONLINEAR ESTIMATION COMPUTATIONS
C                AND CONTROLS OUTPUT SUMMARIES
C
      SUBROUTINE KLANL(C,T,F,R,NOB,CS,CO,XKLA,INAME,NOUT)
      DIMENSION C(100),T(100),F(100),R(100)
      REAL K1,K2,K3
C
C     STEP 1 - WRITE TITLES
C
      WRITE(NOUT,20) INAME
   20 FORMAT(//20X,'NON LINEAR FSTIMATION'/
     115X,'UNSTEADY STATE OXYGEN TRANSFER'/
     223X,'DATA SET',A5)
      WRITE(NOUT,21)
   21 FORMAT(///' ITERATION',29X,'KLA',8X,'SUM OF'/
     12X,'NUMBER',6X,'C-STAR',6X,'C-ZERO',6X,'PRIME',6X,'SQUARES'/)
C
C     STEP 2 - INITIALIZATION OF VARIABLES
C
      K=0
      OSSQ = 0.0
      DO 50 I=1,NOB
      F(I) = CS-(CS-CO)*EXP(-XKLA*T(I))
      R(I) = C(I) - F(I)
   50 OSSQ = OSSQ + R(I)*R(I)
      WRITE(NOUT,30) K,CS,CO,XKLA,OSSQ
   99 K=K+1
      A11=0.0
      A12=0.0
      A13=0.0
      A22=0.0
      A23=0.0
      A33=0.0
      C1=0.0
      C2=0.0
      C3=0.0
      SSQ=0.0
C
C     STEP 3 - SET UP NORMAL EQUATIONS FOR LINEARIZED MODEL
C              USING CURRENT LEAST SQUARES ESTIMATES.
C
      DO 100 I=1,NOB
      Z2=EXP(-XKLA*T(I))
      Z1=1.0-Z2
      Z3 = T(I)*Z2*(CS-CO)
      A11 = A11 + Z1*Z1
      A12 = A12 + Z1*Z2
      A13 = A13 + Z1*Z3
      A22 = A22 + Z2*Z2
      A23 = A23 + Z2*Z3
      A33 = A33 + Z3*Z3
      F(I) = CS - (CS-CO)*Z2
      R(I) = C(I) - F(I)
      C1 = C1 + R(I)*Z1
      C2 = C2 + R(I)*Z2
      C3 = C3 + R(I)*Z3
100   CONTINUE
C
C     STEP 4 - SOLUTION OF NORMAL EQUATIONS FOR CORRECTIONS TO
C              THE PRIOR LEAST SQUARES ESTIMATES
C
      D1 = A11*A22 - A12*A12
      D2 = A11*C3 - A13*C1
      D3 = A11*A23 - A13*A12
      D4 = A33*A11 - A13*A13
      D5 = A11*C2 - A12*C1
      BN3 = D1*D2 - D3 *D5
      BD3 = D1*D4 - D3*D3
      B3 = BN3/BD3
      BN2 = D5 - D3*B3
      B2 = BN2/D1
      B1 = (C1 - A12*B2 - A13*B3)/A11
C
C     STEP 5 - UPDATE ESTIMATES, SUM OF SQUARES
C
      K1 = B1 + CS
      K2 = B2 + CO
      K3 = B3 + XKLA
      DO 120 I=1,NOB
      F(I) = K1 -(K1-K2)*EXP(-K3*T(I))
      R(I) = C(I) - F(I)
120   SSQ = SSQ + R(I)*R(I)
C
```

```
C     STEP 6 - TEST FOR CONVERGENCE - PARAMETERS
C
      IF(B1/K1.GE.0.00001) GO TO 200
      IF(B2/K2.GE.0.00001) GO TO 200
      IF(B3/K3.GE.0.00001) GO TO 200
C
C     ITERATIONS CONVERGED
C
150   WRITE(NOUT,30) K,K1,K2,K3,SSQ
30    FORMAT(3X,I4,3X,4E12.4)
      GO TO 300
C
C     PARAMETERS NOT CONVERGED, TEST SUM OF SQUARES
C
200   IF(ABS((OSSQ-SSQ).LE.0.000001) GO TO 150
C
C     SUM OF SQUARES NOT CONVERGED, TEST ITERATIONS
C
      IF(K.GT.10) GO TO 350
      WRITE(NOUT,30) K,K1,K2,K3,SSQ
C     NEW ESTIMATES
      CS = K1
      CO = K2
      XKLA = K3
      OSSQ = SSQ
      GO TO 99
300   CONTINUE
      XDF = NOB - 3
      RSM = SSQ/XDF
      ERROR = SQRT(RSM)
      WRITE(NOUT,22) ERROR
22    FORMAT(/' ESTIMATE OF ERROR FROM RESIDUAL MEAN SQUARE  ',F6.2)
      WRITE(NOUT,23)
23    FORMAT(//17X,'STANDARD DEVIATIONS'/
     116X,'OF PARAMETER ESTIMATES'//)
C
C     STEP 7 - COMPUTE STANDARD DEVIATIONS OF THE PARAMETER ESTIMATES
C
      DETP = A11*A22*A33 + 2.0*A12*A13*A23
      DETN = A11*A23*A23 + A22*A13*A13 + A33*A12*A12
      DET = DETP-DETN
      E11 = A22*A33 - A23*A23
      E22 = A11*A33 - A13*A13
      E33 = A11*A22 - A12*A12
      VARK1 = (E11/DET)*RSM
      VARK2 = (E22/DET)*RSM
      VARK3 = (E33/DET)*RSM
      SIGCS = SQRT(VARK1)
      SIGCO = SQRT(VARK2)
      SIGKL = SQRT(VARK3)
      WRITE(NOUT,31) SIGCS,SIGCO,SIGKL
```

```
 31     FORMAT(' ABSOLUTE'/3X'UNITS',2X,3E12.4)
        SIGCS = SIGCS/CS*100.0
        SIGCO = SIGCO/CO*100.0
        SIGKL = SIGKL/XKLA*100.0
        WRITE(NOUT,32) SIGCS,SIGCO,SIGKL
 32     FORMAT(/' PER CENT'/2X'OF LSE',2X,3F12.1)
C
C       STEP 8 - WRITE FINAL SUMMARY
C
        WRITE(NOUT,33)
 33     FORMAT(//15X,'SUMMARY OF DATA'//
       110X,'TIME',6X,'CONC',5X,'FITTED',3X,'RESIDUAL'/
       230X,'VALUE'/)
        DO 370 I=1,NOB
 370    WRITE(NOUT,34) I,T(I),C(I),F(I),R(I)
 34     FORMAT(I5,4F10.2)
        GO TO 400
 350    WRITE(NOUT,35)
 35     FORMAT(//3X,'SOLUTION NOT CONVERGED IN 10 ITERATIONS!'//)
 400    RETURN
        END
```

```
FORTRAN:  KLANLN
MAIN.
KLANL
LINK:    Loading
CLNKXCT KLANLN execution

 NAME OF INPUT DATA FILE:  COMB

                          NON LINEAR ESTIMATION
                       UNSTEADY STATE OXYGEN TRANSFER
                              DATA SET COMB
 ITERATION                                        KLA
  NUMBER       C-STAR         C-ZERO             PRIME          SUM OF
                                                               SQUARES
     0      0.1200E+01    0.5000E+00    0.1500E+00    0.4893E+02
     1      0.1116E+02    0.1935E+01    0.6928E-01    0.6109E+01
     2      0.1133E+02    0.1186E+01    0.8733E-01    0.7795E-01
     3      0.1143E+02    0.1122E+01    0.8691E-01    0.1617E-01
     4      0.1143E+02    0.1122E+01    0.8692E-01    0.1617E-01
     5      0.1143E+02    0.1122E+01    0.8692E-01    0.1617E-01
ESTIMATE OF ERROR FORM RESIDUAL MEAN SQUARE                0.03

                            STANDARD DEVIATIONS
                           OF PARAMETER ESTIMATES
ABSOLUTE
  UNITS        0.1822E-01   0.3769E-01   0.6443E-03

PER CENT
 OF LSE              0.2          3.4          0.7

                               SUMMARY OF DATA
               TIME         CONC            FITTED          RESIDUAL
                                            VALUE
 1             2.00          2.77            2.77             0.00
 2             4.00          4.15            4.15             0.00
 3             6.00          5.35            5.31             0.04
 4             8.00          6.25            6.29            -0.04
 5            10.00          7.08            7.11            -0.03
 6            12.00          7.80            7.80             0.00
 7            14.00          8.34            8.37            -0.03
 8            16.00          8.85            8.86            -0.01
 9            18.00          9.28            9.27             0.01
10            20.00          9.62            9.61             0.01
11            22.00          9.93            9.90             0.03
12            25.00         10.24           10.25            -0.01
13            30.00         10.70           10.67             0.03
14            35.00         11.00           10.93             0.07
15            40.00         11.14           11.11             0.03
16            45.00         11.20           11.22            -0.02
17            50.00         11.25           11.29            -0.04
18            55.00         11.30           11.34            -0.04
```

# Appendix B
## Basic Program for Non Linear Regression

This appendix gives the BASIC computer language adaption of the FORTRAN non linear estimation program discussed in Appendix A. Also given are examples of output obtained by applying the BASIC program to typical data sets using an APPLE II Microcomputer.

```
10  D$ =  CHR$ (4)
20   REM
30   REM  NON-LINEAR LEAST SQUARES PROGRAM IN APPLE II BASIC
40   REM  FOR OXYGEN TRANSFER PARAMETERS
50   REM  OUTPUT SETUP FOR 40 POSITION CRT/MONITOR
60   TEXT : CALL  - 936: REM  CLEARS SCREEN
70   REM
80   REM  ::::::::::::::::::::::
90   REM  STEP 1
100   REM  WRITE TITLES
110   REM  ::::::::::::::::::::::
120   REM
130   PRINT "****************************************"
140   PRINT "        NON-LINEAR ESTIMATION FOR"
150   PRINT "      UNSTEADY-STATE OXYGEN TRANSFER"
160   PRINT "****************************************"
170   PRINT "                  BY"
180   PRINT "LINFIELD C. BROWN & GEORGE R. FISETTE"
190   PRINT "VERSION  1.0-NOVEMBER 11, 1979"
200   PRINT
210   INVERSE : PRINT "THE VALUES ARE TRUNCATED": PRINT "AND NOT ROUNDED OFF." :P
RINT  : NORMAL
220   REM
230   REM  PROGRAM HAS MAXIMUM LIMIT OF 30 DATA POINTS
240   REM
250   DIM C(30),T(30),F(30),R(30)
260   INPUT "IS DATA IN DISK FILE;Y/N?";A$
270   INPUT "INPUT NAME OF DATE FILL?";N$
280   IF A$ = " " GOTO 650: REM  GET DATA FROM DISK FILL
290   INPUT "DO YOU WANT INPUT DATA SAVED ON DISK,Y/N?";A$
300   PRINT "INPUT DATA IN TIME,DO DATA PAIRS"
310   PRINT "INPUT 999,999 AS LAST DATA PAIR"
320   FOR I = 1 TO 30
330   INPUT T(I),C(I)
340   IF T(I) = 999.0 GOTO 360
350   NEXT I
360  ND = I - 1.0
370   INPUT "BEST ESTIMATE FOR C-STAR OR USE 10.0 MG/L?";CS
380   INPUT "BEST ESTIMATE FOR C-ZERO OR USE 0.0 MG/L?";CO
390   INPUT "BEST ESTIMATE FOR KLA-PRIME OR USE 4.0 1/HR?";XK
400  XK = XK / 60.0
410   IF A$ = "N" GOTO 790
420   REM
430   REM  WRITE DATA TO DISK FILE
440   REM  SPECIFIC FOR APPLE/MICROSOFT BASIC
450   REM
460   PRINT D$;"OPEN "N$;",VO,L15"
470   FOR I = 1 to ND
480   PRINT D$;"WRITE "N$;",BO,R";I
490   PRINT T(I): PRINT C(I)
500   NEXT I
510   PRINT D$;"WRITE "N$;",BO,RO"
520   PRINT ND
530   PRINT D$;"WRITE "N$;",BO,R";ND + 1.
540   PRINT CS
```

```
550   PRINT D$;"WRITE "N$;",BO,R";ND + 2.
560   PRINT CO
570   PRINT D$;"WRITE "N$;",BO,R";ND + 3.
580   PRINT XK
590   PRINT D$;"CLOSE "N$
600   GOTO 790
610   REM
620   REM  READ DISK FILE FOR DATA
630   REM  SPECIFIC FOR APPLE/MICROSOFT BASIC
640   REM
650   PRINT D$;"OPEN "N$;",VO,L15"
660   PRINT D$;"READ "N$;",BO,RO"
670   INPUT ND
680   FOR I = 1 TO ND
690   PRINT D$;"READ "N$;",BO,R";I
700   INPUT T(I),C(I)
710   NEXT I
720   PRINT D$:"READ "N$;",BO,R";ND + 1.
730   INPUT CS
740   PRINT D$;"READ "N$;",BO,R";ND + 2.
750   INPUT CO
760   PRINT D$;"READ "N$;",BO,R";ND + 3.
770   INPUT XK
780   PRINT D$;"CLOSE "N$
790   PRINT : FLASH : INPUT "HIT RETURN FOR ITERATIONS.";I$: NORMAL
800   CALL    936: PRINT : PRINT "          DATA SET ";N$: PRINT
810   PRINT "ITERATION" TAB( 11)"C-STAR" TAB( 18)"C-ZERO" TAB( 26)"KLA" TAB ( 33)"
SUM OF"
820   PRINT TAB( 2)"NUMBER" TAB( 26)"PRIME" TAB( 33)"SQUARES"
830   PRINT TAB ( 11)"(MG/L)" TAB( 18)"(MG/L)" TAB ( 26)"(1/HR)"
840   PRINT
850   REM
860   REM  ::::::::::::::::::::::
870   REM  STEP 2
880   REM  INITIALIZATION OF VARIABLES
890   REM  DO ITERATION CALCULATIONS
900   REM  ::::::::::::::::::::::
910   REM
920  K% = 0
930  OS = 0.0
940   FOR I = 1 TO ND
950  F(I) = CS - (CS - CO) * EXP ( - XK * T(I))
960  R(I) = C(I) - F(I)
970  OS = OS + R(I) * R(I)
980   NEXT I
990  ZZ$ = STR$ (CS) :VA = 5.: GOSUB 2900
1000  CS$ = ZZ$:ZZ$ = STR$ (CO): GOSUB 2900
1010  CO$ = ZZ$:ZZ$ = STR$ (XK * 60.): GOSUB 2900
1020  XK$ = ZZ$:ZZ$ = STR$ (OS): GOSUB 2900
1030  OS$ = ZZ$
1040   PRINT TAB( 4)K% TAB( 10)CS$ TAB( 18)CO$ TAB( 26)XK$ TAB( 33)OS$
1050   GOTO 1070
1060   REM
1070   REM  CALCULATION LOOP - INITILIZE VARIABLES
```

```
1080  REM
1090 K% = K% + 1
1100 A1 = 0.0
1110 A2 = 0.0
1120 A3 = 0.0
1130 A4 = 0.0
1140 A5 = 0.0
1150 A6 = 0.0
1160 C1 = 0.0
1170 C2 = 0.0
1180 C3 = 0.0
1190 SQ = 0.0
1200  REM
1210  REM  ::::::::::::::::::::::
1220  REM  STEP 3
1230  REM  SETUP NORMAL EQUATIONS FOR LINEARIZED MODEL
1240  REM  USING CURRENT LEAST SQUARE ESTIMATES
1250  REM  ::::::::::::::::::::::
1260  REM
1270  FOR I = 1 TO ND
1280 Z2 = EXP ( - XK * T(I))
1290 Z1 = 1.0 - Z2
1300 Z3 = T(I) * Z2 * (CS - CO)
1310 A1 = A1 + Z1 * Z1
1320 A2 = A2 + Z1 * Z2
1330 A3 = A3 + Z1 * Z3
1340 A4 = A4 + Z2 * Z2
1350 A5 = A5 + Z2 * Z3
1360 A6 = A6 + Z3 * Z3
1370 F(I) = CS - (CS - CO) * Z2
1380 R(I) = C(I) - F(I)
1390 C1 = C1 + R(I) * Z1
1400 C2 = C2 + R(I) * Z2
1410 C3 = C3 + R(I) * Z3
1420  NEXT I
1430  REM
1440  REM  ::::::::::::::::::::::
1450  REM  STEP 4
1460  REM  SOLUTION OF NORMAL EQUATIONS FOR CORRECTIONS
1470  REM  TO THE PRIOR LEAST SQUARES ESTIMATES
1480  REM  ::::::::::::::::::::::
1490  REM
1500 D1 = A1 * A4 - A2 * A2
1510 D2 = A1 * C3 - A3 * C1
1520 D3 = A1 * A5 - A3 * A2
1530 D4 = A6 * A1 - A3 * A3
1540 D5 = A1 * C2 - A2 * C1
1550 XN = D1 * D2 - D3 * D5
1560 XD = D1 * D4 - D3 * D3
1570 X3 = XN / XD
1580 YN = D5 - D3 * X3
1590 X2 = YN / D1
1600 X1 = (C1 - A2 * X2 - A3 * X3) / A1
1610  REM
```

```
1620  REM  ::::::::::::::::::::::
1630  REM  STEP 5
1640  REM  UPDATE ESTIMATES, SUM OF SQUARES
1650  REM  ::::::::::::::::::::::
1660  REM
1670  T1 = X1 + CS
1680  T2 = X2 + CO
1690  T3 = X3 + XK
1700   FOR I = 1 TO ND
1710  F(I) = T1 - (T1 - T2) * EXP ( - T3 * T(I))
1720  R(I) = C(I) - F(I)
1730  SQ = SQ + R(I) * R(I)
1740   NEXT I
1750   REM
1760   REM  ::::::::::::::::::::::
1770   REM  STEP 6
1780   REM  TEST FOR CONVERGENCE - PARAMETERS   1 PART IN 100,000
1790   REM  ::::::::::::::::::::::
1800   REM
1810   IF (X1 / T1 ≥ 0.00001) AND (X2 / T2 ≥ 0.00001) AND (X3 / T3 ≥ 0.00001) GOT
0 2160
1820   REM
1830   REM  PARAMETERS NOT CONVERGED,
1840   REM  TEST SUM OF SQUARES - 1 PART IN 1,000,000
1850   REM
1860   IF ABS ((OS - SQ)/SQ)≤0.000001 GOTO 2160
1870   REM
1880   REM  SUM OF SQUARES NOT CONVERGED,
1890   REM  TEST NO. OF ITERATIONS
1900   REM
1910   IF (K%   10) GOTO 2090
1920  ZZ$ = STR$ (T1) : GOSUB 2900
1930  T1$ = ZZ$:ZZ$ =  STR$ (T2) :  GOSUB 2900
1940  T2$ = ZZ$:ZZ$ =  STR$ (T3  * 60.): GOSUB 2900
1950  T3$ = ZZ$:ZZ$ =  STR$ (SQ) : GOSUB 2900
1960  SQ$ = ZZ$
1970   PRINT TAB( 4)K% TAB( 10)T1$ TAB( 18)T2$ TAB( 26)T3$ TAB( 33)SQ$
1980   REM
1990   REM  NEW ESTIMATES
2000   REM
2010  CS = T1
2020  CO = T2
2030  XK = T3
2040  OS = SQ
2050   GOTO 1090
2060   REM
2070   REM  OUTPUTS
2080   REM
2090   PRINT
2100   PRINT "SOLUTION NOT CONVERGED IN 10 ITERATIONS'"
2110   PRINT  "CHANGE VALUE IN LINE 2670 TO TRY MORE  ITERATIONS."
2120   END
2130   REM
2140   REM  OUTPUT PARAMETER ESTIMATES
```

```
2150  REM
2160 ZZ$ =  STR$ (T1): GOSUB 2900
2170 T1$ = ZZ$:ZZ$ = STR$ (T2): GOSUB 2900
2180 T2$ = ZZ$:ZZ$ = STR$ (T3 * 60.): SOSUB 2900
2190 T3$ = ZZ$:ZZ$ = STR$ (SQ): GOSUB 2900
2200 SQ$ = ZZ$
2210  PRINT TAB( 4)K% TAB( 10)T1$ TAB( 18)T2$ TAB( 26)T3$ TAB( 33)SQ$
2220  PRINT
2230  REM
2240  REM  ::::::::::::::::::::::
2250  REM  STEP 7
2260  REM   COMPUTE STANDARD DEVIATIONS OF PARAMETER ESTIMATES
2270  REM  ::::::::::::::::::::::
2280  REM
2290 XF = ND - 3.0
2300 RS = SQ / XF
2310 ER = SQR (RS)
2320  PRINT "STD DEVIATIONS OF PARAMETER ESTIMATES"
2330  PRINT
2340 DP = A1 * A4 * A6 + 2.0 * A2 * A3 * A5
2350 DN = A1 * A5 * A5 + A4 * A3 * A3 + A6 * A2 * A2
2360 DT = DP - DN
2370 F1 = A4 * A6 - A5 * A5
2380 F2 = A1 * A6 - A3 * A3
2390 F3 = A1 * A4 - A2 * A2
2400 V1 = (F1 / DT) * RS
2410 V2 = (F2 / DT) * RS
2420 V3 = (F3 / DT) * RS
2430 S1 = SQR (V1)
2440 S2 = SQR (V2)
2450 S3 = SQR (V3)
2460 ZZ$ = STR$ (S1):VA = 5.: GOSUB 2900
2470 S1$ = ZZ$:ZZ$ = STR$ (S2): GOSUB 2900
2480 S2$ = ZZ$:ZZ$ = STR$ (S3 * 60.): GOSUB 2900
2490 S3$ = ZZ$
2500  PRINT " UNITS" TAB( 10)S1$ TAB( 18)S2$ TAB( 26)S3$
2510 S1 = S1 / CS * 100.0
2520 S2 = S2 / CO * 100.0
2530 S3 = S3 / XK * 100.0
2540 ZZ$ = STR$ (S1):VA = 3.: GOSUB 2900
2550 S1$ = ZZ$:ZZ$ = STR$ (S2): GOSUB 2900
2560 S2$ = ZZ$:ZZ$ = STR$ (s3): GOSUB 2900
2570 S3$ = ZZ$
2580  PRINT "% OF LSE" TAB( 10)S1$ TAB( 18)S2$ TAB( 26)S3$
2590  PRINT
2600 ZZ$ = STR$ (ER):VA = 4.: GOSUB 2900
2610 ER$ = ZZ$
2620  PRINT "ESTIMATE OF ERROR = ";ER$
2630  REM
2640  REM  ::::::::::::::::::::::
2650  REM  STEP 8
2660  REM  WRITE SUMMARY
2670  REM  ::::::::::::::::::::::
```

```
2680   REM
2690   PRINT
2700   FLASH : INPUT "HIT RETURN FOR SUMMARY OF DATA.";I$: NORMAL
2710   CALL - 936: PRINT : PRINT : REM  CLEARS SCREEN
2720   PRINT TAB( 13)"SUMMARY OF DATA"
2730   PRINT : PRINT
2740   PRINT TAB( 8)"TIME" TAB( 16)"CONC" TAB( 22)"FIT VALUE" TAB( 32)"RESIDUAL"
2750   PRINT TAB( 8)"(MIN)" TAB( 15)"(MG/L)" TAB( 23)"MG/L"
2760   PRINT
2770   FOR I = 1 TO ND
2780  ZZ$ = STR$ (F(I)):VA = 4.: GOSUB 2900
2790  H1$ = ZZ$:ZZ$ = STR$ (R(I)): GOSUB 2900
2800  H2$ = ZZ$
2810   PRINT TAB( 2)I TAB( 8)T(I) TAB( 16)C(I) TAB( 25)H1$ TAB( 33)H2$
2820   NEXT I
2830   PRINT : PRINT
2840   PRINT "****************************************"
2850   END
2860   REM
2870   REM  OUTPUT FORMATTING ROUTINES
2880   REM  SPECIFIC FOR APPLE/MICROSOFT BASIC
2890   REM
2900  LL = LEN (ZZ$)
2910   IF LL < 12 THEN ZZ$ = LEFT$ (ZZ$,VA): RETURN
2920   IF  MID$ (ZZ$,LL - 2,1) = "+" THEN ZZ$ = LEFT$ (ZZ$,VA - 3) + RIGHT$ (ZZ
$ 3):  RETURN
2930  CC = 2.: IF LEFT$ (ZZ$,1)   =   "-" THEN CC = 1.
2940   IF MID$ (ZZ$,LL - 3,1) = "E" THEN EE = VAL (RIGHT$ (ZZ$,2)):NN$ = MID$
 (ZZ$,CC,1): FOR J = 1 TO EE:NN$ = "0" + NN$:: NEXT J:ZZ$ = "." NN$ + MID$ (Z
Z$,CC + 2,LL - 4): IF CC = 2. THEN ZZ$ = "-" + ZZ$
2950  ZZ$ = LEFT$ (ZZ$,VA): RETURN
2960   REM
2970   REM  NON-LINEAR LEAST SQUARES PROGRAM FOR
2980   REM  UNSTEADY-STATE OXYGEN TRANSFER
2990   REM  LY LINFIELD C. BROWN & GEORGE R. FISETTE
3000   REM  VERSION  1.0-NOVEMBER 11, 1979
3010   REM  COPYRIGHT BY ASCE
```

```
RUN
******************************************************************************
          NON-LINEAR ESTIMATION FOR UNSTEADY-STATE OXYGEN TRANSFER
******************************************************************************
BY
LINFIELD C. BROWN & GEORGE R. FISETTE
VERSION  1.0-NOVEMBER 11, 1979

IS DATA IN DISK FILE; Y/NPY
INPUT NAME OF DATA FILE?TEST2

                              DATA SET TEST2
ITERATION
NUMBER       C-STAR           C-ZERO          KLA             SUM OF
                                              PRIME           SQUARES
0            10               .1              .07             .869733335
1            10.1592458       .713872799      .0601645437     .189522011
2            10.1889055       .694460338      .0603440105     .185289531
3            10.1893432       .694630637      .0603354201     .185289365

STANDARD DEVIATIONS OF PARAMETER ESTIMATES

ABSOLUTE UNITS  .195594007    .111803446      3.18746727E-03
PERCENT OF LSE  1.91967633    16.0993277      5.28216013

ESTIMATE OF ERROR = .152187945

SUMMARY OF DATA

             TIME             CONC            FIT VALUE       @ RESIDUAL
1            .77              1.25            1.12564889       .124351112
2            1.75             1.75            1.64603691       .103963087
3            2.67             2.05            2.10734213      -.0573421288
4            5.08             3               3.20108643      -.201086428
5            7.77             4.1             4.24803791      -.148037907
6            10.65            5.25            5.19571448       .0542855244
7            13.3             5.85            5.93358056      -.0835805573
8            17.88            7.1             6.96110108       .138898922
9            23.4             8.1             7.87555468       .224445321
10           34.52            8.9             9.00645823      -.106458228
11           49.13            9.65            9.69943839      -.0494383909

******************************************************************************
```

# OXYGEN TRANSFER TESTING UNDER PROCESS CONDITIONS

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## INTRODUCTION

Adequate test procedures must be available to permit the investigator to assess aeration equipment performance during actual process operation. Several test methods have been employed to determine oxygen transfer under field conditions.

In general, the methods can be categorized according to the rate of change of dissolved oxygen (DO) in a given reactor (or segment of reactor). Systems in which the rate of DO change is zero at any given point are referred to as steady state systems; the others are classified as non-steady state systems. In some cases, the influent wastewater may be diverted from a reactor being tested. These are referred to as batch tests. The term "continuous test" is used for those cases where the influent wastewater flow is not diverted.

Several _in situ_ test methods do not require a direct measure of the oxygen-uptake rate. These have been broadly categorized as the mass balance method, the off-gas method, and the tracer method. The mass balance method requires data on the net change in oxidation level between all entering and exiting liquid flows. The off-gas method is simply a mass balance on oxygen that includes both the liquid and gas streams. The tracer method indirectly measures the rate of oxygen transfer by determining the rate of transfer of a radioactive tracer.

The most commonly used methods involve the direct measurement of the oxygen uptake rate, R, of a respiring biological system. Two of these methods are carried out with little or no DO variation in either batch or continuous flow systems; these are referred to as "Steady State Batch Tests" and "Steady State Continuous Tests," respectively. In the remaining two tests, the DO level in the reactor is adjusted at the beginning of the test to be either greater than or less than the steady state DO. These tests are referred to as "Non-steady State Batch Tests" if the influent wastewater flow is discontinued for the test and "Non-steady State Continuous Tests" if the influent wastewater flow is continued during testing.

## KEY FACTORS IN CONVENTIONAL TESTING

The successful evaluation of aeration equipment _in situ_ using conventional techniques involves careful measurements of several key parameters under full-scale operating conditions. A discussion of these measurements and their importance in field evaluations follows.

## Dissolved Oxygen

An accurate measure of the DO concentration in an aeration volume is essential to any evaluation of aeration equipment. For in situ testing, direct-reading DO probes are the only practical means of measuring the DO concentration in mixed liquor suspended solids (MLSS) samples. Following proper calibration of the probes, considerable care and attention are required to assure continuous reliable results under field conditions.

Aeration testing may be carried out on an entire tank or on an isolated mixing zone within a test volume. Typically, a minimum of three probes are used in field testing; and, depending on the type of aeration system being tested, placement of the units in the test tank can be critical. For example, in aeration volumes with a relatively uniform DO concentration (eg, well mixed with respect to DO), probes may be located in the tank without particular regard to aerator mixing pattern. However, for aeration volumes that are not well mixed with respect to DO, probes should be strategically placed around the flow pattern characteristic of the aeration device. Depending on the specific application, many aeration devices will not yield a uniform DO concentration in the tank under test conditions. Thus, it is important to establish a complete DO profile on these aeration systems prior to testing. The DO profile will indicate the actual mixing characteristics of the test tank and allow proper placement of the DO probes.

During field testing of aeration systems, the DO concentration should not be the limiting factor in the biological reaction. For instance, the DO can become limiting at approximately 0.5 mg/ℓ for non-nitrifying activated sludge systems, and at approximately 1.0 to 1.5 mg/ℓ for nitrifying activated sludge systems. Therefore, testing where the minimum DO is above 0.5 mg/ℓ and 1.5 mg/ℓ for non-nitrifying and nitrifying activated sludge systems, respectively, is essential.

## Oxygen Uptake Rate

A significant factor for evaluating aeration equipment under operating conditions is an accurate measurement of the MLSS oxygen uptake rate, R. Experience has shown that accurate measurement of the rapid oxygen uptakes created by high organic loads is virtually impossible. Ideally, the oxygen uptake rate measurements should be taken in situ or immediately at the point of sample collection. However, as a practical matter, a finite time period elapses prior to field measurements of this parameter. Since the oxygen uptake rate of a sample will vary as the available soluble substrate is oxidized, significant variations in the uptake rate may be observed for samples taken from moderately to highly loaded systems. Thus, caution is urged where oxygen uptake rates are being measured under actual plant loading situations.

A practical approach used to minimize the variability of the oxygen uptake rate is testing under endogenous respiration conditions. Endogenous respiration can be achieved by diverting the influent wastewater to other aeration tanks in order to allow the MLSS to assimilate the remaining soluble organics. This procedure is the basis for batch testing and establishes a low, relatively constant oxygen uptake rate that can be more accurately measured.

## Alpha and Beta Factors

Alpha ($\alpha$) and beta ($\beta$) values are necessary for describing the influence of mixed liquor characteristics (eg, dissolved substances and suspended solids) on the transfer capability of the aeration equipment in clean water. Generally, $\alpha$ and $\beta$ measurements are made only when the field transfer rates are to be compared to standardized transfer rates developed under clean water conditions.

Due to the variability of influent wastewater quality, mixed liquor characteristics, and type of aeration device, the $\alpha$ level may be extremely variable for a given application. Testing under endogenous respiration conditions is advantageous because it tends to minimize the variability of $\alpha$ during the test period. However, the oxygen transfer rate obtained cannot be assumed to be the same as that occurring during operating conditions under normal plant loading. A more detailed review of the complexities associated with $\alpha$ and $\beta$ factors has been presented by Stenstrom and Gilbert (1).

## Wastewater Temperature

Wastewater temperature also affects the evaluation of field oxygen transfer rates. Both oxygen transfer rate and oxygen update rate are significantly altered by changes in temperature. Biological reactions are thought to be governed by a doubling of rate for every 10°C increase in temperature [corresponds to $R_T = R_{20}\ 1.072^{(T-20)}$] within the practical operating range of 10-30°C (2). Thus, if an oxygen uptake rate is measured at a wastewater temperature different from the actual operating temperature, the appropriate temperature correction for reaction rate must be made during data evaluation.

At the same time, since aeration equipment is typically rated at 20°C, but is usually tested at some other temperature, appropriate transfer rate adjustments must also be applied to account for the effects of wastewater temperature. The temperature correction for oxygen transfer has been discussed in detail by Stenstrom and Gilbert (1). Where practical, testing at temperatures near 20°C is desirable, since it will minimize temperature corrections and allow a more accurate comparison with transfer rates at standard conditions.

# REVIEW OF TEST METHODS

## Steady State Testing

### Test Descriptions

Steady state testing involves simultaneous measurement of DO and oxygen uptake rates in full-scale aeration tanks. Testing may be conducted with or without influent wastewater flow to the aeration tank. For convenience, these testing methods are referred to as the continuous test (with wastewater flow) and the batch endogenous test (without wastewater flow).

The continuous test is the classical approach of determining field oxygen transfer in an aeration tank under normal operating conditions (3-6). Generally, the DO, oxygen uptake rate, and temperature are monitored in the tank as wastewater continues to flow through during a specified test period. The DO of the influent and effluent streams are also measured throughout the test. The batch endogenous test is conducted by discontinuing the influent wastewater flow to the aeration tank prior to the test period. The recycle sludge flow to the aeration tank may be held constant or discontinued during the test period. Operation in the endogenous phase should yield more uniform aeration tank contents (eg, R, DO, and $\alpha$ value) and should result in increased testing accuracy.

Data Evaluation

The data obtained from the steady state tests may be evaluated using Equation 1 as described below:

$$K_La_f = \frac{\underline{\text{System Test Constant}}}{C^*_{\infty f} - C_R} \qquad (1)$$

The system test constant is defined as follows for batch and continuous testing:

Batch Test

$$B_1 = R - \frac{Q_r}{V}(C_r - C_R) \text{ (with recycle sludge flow)} \qquad (2)$$

$$B_1 = R \text{ (without recycle sludge flow)} \qquad (3)$$

Continuous Test

$$F_1 = R - \frac{Q'}{V}(C'_o - C_R) \qquad (4)$$

where:

$C^*_{\infty f}$ = saturation DO concentration, mg/ℓ in mixed liquor

$C_R$ = DO level, mg/ℓ, that most accurately represents the driving force, $C^*_{\infty f} - C_R$, in the transfer zone of the system.

$C_r$ = DO level, mg/ℓ, in the recycle sludge flow, $Q_r$

$C'_o$ = DO level, mg/ℓ, in the test volume influent flow, $Q'$

R = average oxygen uptake rate, mg/ℓ/hr, of the MLSS in the test volume, V

$Q_r$ = recycle sludge flow, ℓ/hr, to the test volume, V

Q' = influent flow, ℓ/hr, to the test volume, V

V = test volume, ℓ

The resulting $K_La_f$ represents the field oxygen transfer rate for the specified test temperature.

Test Limitations

The limitations of a specific test method must be recognized to ensure adequate data collection and allow an appreciation for the accuracy of the end result. For the steady state testing approaches, the following discussion outlines the specific test limitations that have been identified.

Continuous testing - An accurate determination of the value of R may be very difficult to obtain if system influent wastewater flow and substrate concentration are not nearly constant during the test period (6, 7). In addition, if the system being tested has a high organic loading, the value of R will be practically impossible to determine accurately. Thus, the accuracy of the $K_La_f$ determination is directly tied to the ability to determine a representative value of R for the system. McKinney and Stukenberg (8) found where R values were greater than 60 mg/ℓ/hr errors in measurements can be significant and increase with increasing R values.

The inability to control the $\alpha$ value of the test volume during the test period may be a serious drawback to continuous testing. The $\alpha$ level may vary substantially (as much as ±20 percent) at a given location within the tank during the test duration due to variable influent wastewater characteristics. In addition, the $\alpha$ level may even vary spatially within the tank at a given time during the test depending on the degree of treatment and system mixing patterns.

For many applications, specifically for surface aeration installations, the DO concentration may vary spatially within the tank at a given time during testing under continuous loading conditions. Therefore, the continuous testing approach may not be valid for those applications where such variation exists. The requirement of maintaining the excess DO level, $C_R$, within a practical range during the test period may also be a limitation for some test applications. In order to maintain a reasonable range for accurate DO determination, the measured $C_R$ level should be between a minimum of 2 mg/ℓ and a maximum of 75 percent of the system $C^*_{\infty f}$. Limiting the test conditions to within this range will ensure the statistical validity of the DO measurement and provide a more meaningful end result.

The inability to determine a $C^*_{\infty f}$ value for submerged aeration systems effectively limits the use of the continuous test. This technique may be used for submerged aeration systems only where it is possible to perform the test at a minimum of three different oxygen uptake rates. Different oxygen uptake rates may be established by varying the MLSS concentration under a constant organic loading rate or by varying the organic loading rate for a constant MLSS concentration.

Batch testing - To achieve an endogenous respiration condition, the influent flow to the test volume must be discontinued. For systems with limited operational flexibility, this approach may not be pracitcal.

When testing under endogenous repiration conditions, the $\alpha$ level would not necessarily reflect the $\alpha$ level at normal operating conditions. Therefore, if a field transfer rate under operating conditions is desired, the batch test may not yield an accurate estimation of that value. Depending on the specific application, testing under endogenous conditions would be expected to yield higher field transfer rates due to higher $\alpha$ levels. On the other hand, if the test is performed for comparison with a standard oxygen transfer rate, this approach should provide representative information for that comparison.

The requirement of maintaining the excess DO level, $C_R$, within a practical range during the test period may prove to be a limitation for some applications of steady state batch testing as with continuous testing. Also, the inability to determine a $C^*_{\infty f}$ value for submerged aeration systems effectively further restricts the use of the batch test approach (see discussion in previous subsection on continuous test limitations).

## Non-Steady State Testing

### Test Descriptions

Non-steady state testing involves reduction or discontinuation of aeration in the test tank. This allows biological action, through the oxygen uptake of the MLSS, to reduce the DO concentration in the test volume. After the DO level has been depressed sufficiently, aeration is increased or reintroduced to the test tank. By simultaneously monitoring the increase in DO level and the oxygen uptake rate, the field oxygen transfer rate may be determined using a classical reaeration approach. Testing may be conducted with or without influent flow to the aeration tank. As with steady state testing, these two methods are referred to as the continuous test (with wastewater flow) and the batch endogenous test (without wastewater flow).

A batch endogenous desorption test that is very similar to the non-steady state batch endogenous test has been proposed by Kayser (9). Rather than starting the test at a very low DO concentration, the desorption test starts at a supersaturated DO level in the test solution. The premise of the test is that the rate of oxygen desorption is the same as the rate of oxygen transfer to the water. The initial high DO concentration is achieved by a slug addition of hydrogen peroxide uniformly distributed over the entire test volume. Following the initial peroxide addition, by simultaneously monitoring the decrease in the DO level and the oxygen uptake rate at the same test tank locations, the field oxygen transfer rate may be determined using a classical deaeration approach.

**Data Evaluation**

The data obtained from the non-steady state test may be evaluated using Equation 5 as described below:

$$(C_R - C) = (C_R - C_o)\ \exp- \left[(K_L a_f + \text{System Time Constant})t\right] \qquad (5)$$

where:

$C$ = DO concentration, mg/ℓ, at any time, t

$C_R$ = DO level, mg/ℓ, that most accurately represents the driving force, $C^*_{\infty f} - C_R$, in the transfer zone of the system

$C_o$ = DO concentration at test time, t = 0, mg/ℓ

System Time Constant is:

Batch Test

$\frac{Q_r}{V}$ (with recycle sludge flow)

0 (without recycle sludge flow)

Continuous Test

$$\frac{Q'}{V}$$

Equation 5 is the same general form as the equation used for clean water evaluations. Therefore, the same evaluation techniques are applicable, except that for respiring systems the term ($K_L a_f$ + System Time Constant) is obtained where $K_L a$ would be determined for clean water testing.

Once $K_L a_f$ has been determined, the corresponding effective field DO saturation concentration, $C^*_{\infty f}$, can be calculated. This calculation may be performed utilizing Equation 6 as follows:

$$C^*_{\infty f} = C_R + \frac{R + (\text{System Time Constant})(C_R - C'_o)}{K_L a_f} \qquad (6)$$

where:

$C'_o$ = DO level, mg/ℓ, in the combined wastewater influent and sludge recycle flow, Q', for continuous testing and DO level in the sludge recycle flow, $Q_r$, for batch testing

## Test Limitations

For the non-steady state testing approaches, the following discussion outlines the specific test limitations that have been identified.

Continuous testing - An accurate determination of the R value may be very difficult to obtain if influent flow and substrate concentration vary substantially during the test period. If either of these conditions exist, the reaeration curve may be erratic and the calculated $C^*_{\infty f}$ value, which is calculated using R, may be in error. As previously mentioned, a high organic loading to the test system ($R > 60$ mg/ℓ/hr) will also preclude an accurate measurement of R.

Due to the flow-through testing conditions, the influent of non-uniform influent wastewater flow and substrate concentration will cause variable system R and $\alpha$ values over the test period. In addition, under flow-through conditions, the maldistribution of the influent flow throughout the tank volume can bring about spatial variation of R and $\alpha$ values within the test volume. Therefore, testing should be conducted during the operational periods when fluctuations in influent flow and organic strength are minimal.

For many applications, specifically for surface aeration installations, the DO concentration may vary spatially within the tank at a given time during testing under continuous loading conditions. Therefore, this test approach may not be valid for those applications where such variation exists.

For systems with a low oxygen transfer rate and a high organic loading rate, the system equilibrium DO level, $C_R$, may be relatively low. To obtain sufficient test information for statistical data analysis, a reasonable range for the test parameter ($C_R - C$) is essential. Since the practical initial DO level will be approximately 0.5 mg/ℓ, a minimum $C_R$ level of 4.0 mg/ℓ is recommended for reliable evaluation of test data. This provision may prohibit continuous testing at some installations.

For surface aeration installations, the only method of reducing the DO concentration sufficiently may be to completely shut off the aerators or to operate them intermittently. This action would most likely result in inadequate mixing of the test volume, bringing about solids segregation during the deaeration phase of the testing. As the units are turned on the the reaeration phase of the testing, erratic results will be produced during the initial data collection period due to the re-establishment of the uniform mixing regime. Depending on the size and geometric configuration of the test volume, the effect of inadequate mixing could lead to erroneous test results.

Batch testing - This method has the same basic limitations as the steady state batch approach. Additionally, for surface aeration installations, the only method of reducing the DO concentration sufficiently may be to completely shut off the aerators or operate them intermittently. This action would result in inadequate mixing of the test volume and cause erratic results as described for non-steady state continuous testing.

Batch desorption testing - This method has the same basic limitations as the conventional non-steady state batch method. In addition, this method is based upon the assumption that the hydrogen peroxide immediately and fully disassociates to DO and water when added to the test volume. Furthermore, this approach is based on the premise that the rate of oxygen desorption from water-to-air is the same rate as oxygen transfer from air-to-water. Kayser's work (10) seems to verify both of these assumptions. However, if the peroxide does not fully disassociate before the start of the test, the resulting oxygen transfer rate determination will be erroneously lower than the actual field transfer rate.

## Other Test Approaches

In addition to the above tests, other procedures have also been employed to evaluate aeration equipment operating in biological treatment systems. For the most part, these procedures have not been extensively used and, therefore, are not primary tests for respiring system field tests. However, these tests may have application for specific treatment operations.

### Mass Balance: Activated Sludge Systems

The mass balance approach has been proposed to determine oxygen transfer in operational activated sludge systems. Total oxygen balances must be made on the influent, effluent, and waste activated sludge flows. The change in total oxygen across the entire activated sludge system equals the oxygen transferred by the aeration system. Total oxygen measurements should be based on COD, with correction for nitrification. The major problems with the mass balance technique lie in satisfactory measurement of waste activated sludge volumes and in obtaining representative samples of waste activated sludge for analysis. The high suspended solids concentrations in waste activated sludge make it difficult to obtain valid data because of large errors that can result from minor variations in the aeration volume solids inventory. While is is possible to achieve accurate mass balance measurements on small laboratory systems, this technique has limited practical value in field-scale evaluation of aeration equipment for activated sludge systems.

### Mass Balance: Aerated Stabilization Basin Systems

The mass balance procedure has been applied to low-rate aerated stabilization basin (ASB) systems because oxygen transfer cannot be readily measured using more direct methods. The accuracy of this method depends on the extent to which all factors that supply and withdraw oxygen from the system are measured. A number of these factors can be minimized during the test period so that the main factor that determines the aerator oxygen transfer is the reduction in BOD through the respiring system. The mass balance method is simple and straightforward, especially when the ASB system is aerator (oxygen) limited and where significant settleable biological solids are not produced (11).

### Off-Gas Analysis

Off-gas analysis has been used with full-scale, diffused aeration activated sludge systems and covered tank, pure oxygen activated sludge systems (12, 13). A mass balance on oxygen in the gas phase is required; therefore, this method is not applicable to surface aeration equipment. For systems that are not enclosed, a collector or hood is placed in the zone of interest in the aeration tank so a representative sample can be collected and analyzed for residual oxygen. As with the other methods, the validity of this technique depends on the accuracy and precision of the various parameters that have to be measured. For low efficiency of oxygen transfer (3 to 8 percent) associated with many diffused aeration systems, obtaining the requisite accuracy and precision of the mass balance on the input and discharge gases is difficult under the best of conditions. However, it is recognized that significant advances have been made in the recent past regarding instruments and techniques for measuring gas-phase oxygen concentration. Furthermore, aeration systems with oxygen transfer efficiencies of about 10 percent or greater are becoming more the rule than the exception. Therefore, the combined effect of these developments suggests the desirability of additional investigation into the utility and merit of this method.

### Tracer Method

A radioactive tracer technique has been proposed to measure oxygen transfer rate in any aeration system, either in clean water or in wastewater (14). The tracer method requires considerable planning, special radioactive counting equipment, and a radioactive materials license to use the radioisotopes, krypton-85, and tritium. The basic concept of the radioactive tracer technique involves direct measurement of the mass transfer of krypton-85, which is related to the oxygen transfer rate. The tritium is used to measure the dispersion of the tracers in the aeration tank. Both tracers are added to the aeration tank at a single point. As the tracers are dispersed in the aeration tank, the conservative tracer, tritium, mixes with the mixed liquor while the krypton-85 is mixed and stripped off in the gas phase. The key to this procedure lies in the fact that the krypton-85 stripping rate from the mixed liquor is directly related to the oxygen transfer rate from the gas into the mixed liquor. In effect, the tracer method is an indirect method for measuring oxygen transfer. A series of grab samples are collected and counted in a scintillation counter. Care must be taken to insure that gas bubbles to not form in the samples as the krypton-85 would come to equilibrium with the gas bubbles and produce an error in the radioactive counts. Accurate counting of the samples is essential for obtaining good results. The counting efficiency is about 30 percent for tritium and about 90 percent for krypton-85. The tracer technique has had limited application to date but has considerable potential in the oxygen transfer measurement field. More studies are urged using this technique.

## FIELD TEST RESULTS

### Test Conditions

Since both the off-gas analysis and tracer method have had limited use but show excellent potential, a field study was undertaken utilizing both techniques in parallel. The purpose of the test was to determine the field oxygen transfer rate using each test method in a full scale biological treatment system and compare results.

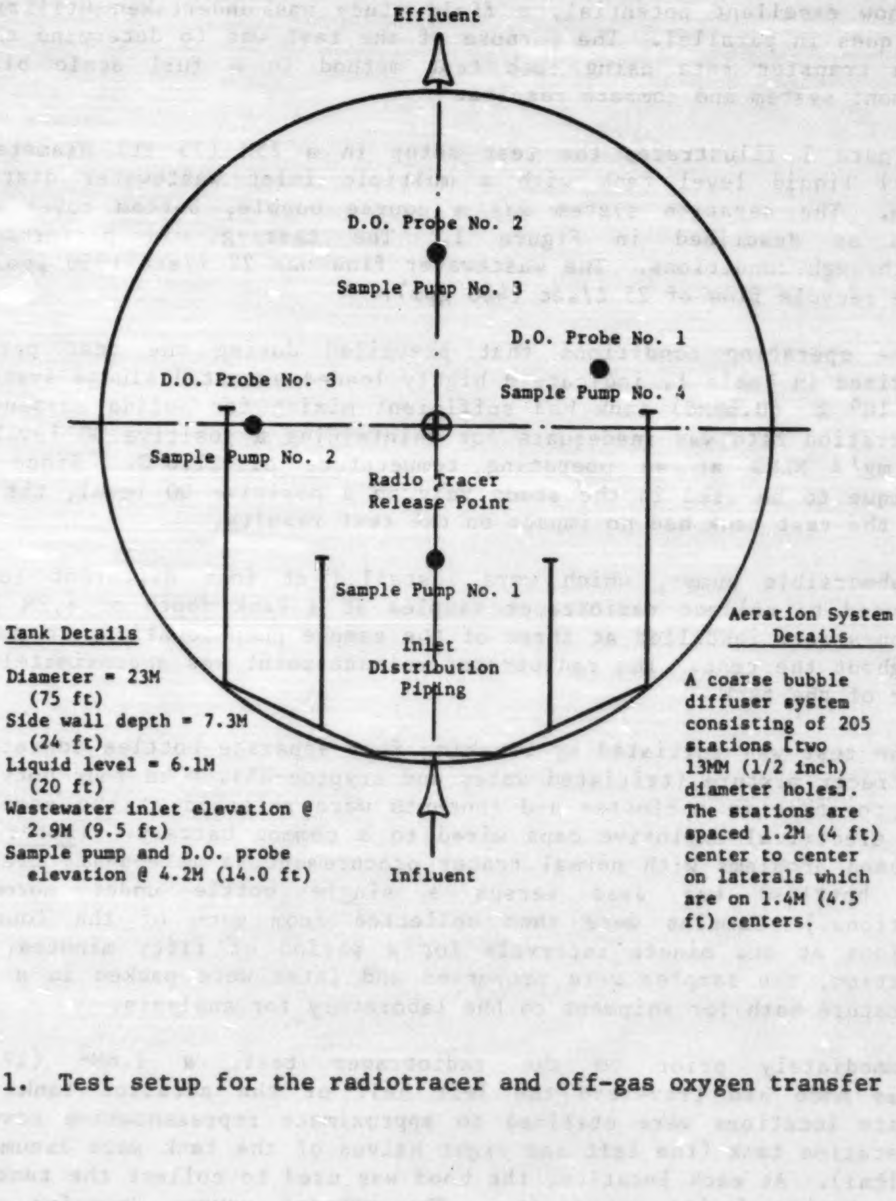
Figure 1 illustrates the test setup in a 23M (75 ft) diameter, 6.1M (20 ft) liquid level tank with a multiple inlet wastewater distribution system. The aeration system was a course bubble, bottom cover diffuser system as described in Figure 1. The testing was performed under flow-through conditions. The wastewater flow was 22 ℓ/sec (350 gpm) with a sludge recycle flow of 25 ℓ/sec (400 gpm).

The operating conditions that prevailed during the test period, as summarized in Table 1, indicate a highly loaded actuated sludge system. The $2.5 \times 10^6$ ℓ (0.66MG) tank had sufficient mixing for solids suspension but the aeration rate was inadequate for maintaining a positive DO level in the 6100 mg/ℓ MLSS at an operating temperature of 21.8°C. Since neither technique to be used in the study rely on a positive DO level, the lack of DO in the test tank had no impact on the test results.

Submersible pumps, which were installed at four different locations, were used to collect radiotracer samples at a tank depth of 4.2M (14 ft). DO probes were installed at three of the sample pump locations to monitor DO throughout the test. The radiotracer release point was approximately in the center of the tank.

The test was initiated by breaking four separate bottles containing the radiotracer mixture (tritiated water and krypton-85). The four bottles were wired together in a cluster and contents were released at the same instant using electrical explosive caps wired to a common battery. [Note: Due to unforseen problems with normal tracer procurement, a make-shift tracer unit (four bottles) was used versus a single bottle under normal test conditions.] Samples were then collected from each of the four sample locations at one minute intervals for a period of fifty minutes. During collection, the samples were preserved and later were packed in a constant temperature bath for shipment to the laboratory for analysis.

Immediately prior to the radiotracer test, a $1.6M^2$ (17.5 $ft^2$) off-gas hood had traversed the left half of the aeration tank. Twelve separate locations were utilized to approximate representative coverage of the aeration tank (the left and right halves of the tank were assumed to be identical). At each location, the hood was used to collect the tank off-gas for analysis of its components. The percent oxygen transfer could be calculated by comparing the hood collected oxygen concentration with that in background air.


Effluent
D.O. Probe No. 2
Sample Pump No. 3
D.O. Probe No. 1
Sample Pump No. 4
D.O. Probe No. 3
Sample Pump No. 2
Radio Tracer
Release Point
Sample Pump No. 1
Inlet
Distribution
Piping
Influent

**Tank Details**

Diameter = 23M (75 ft)
Side wall depth = 7.3M (24 ft)
Liquid level = 6.1M (20 ft)
Wastewater inlet elevation @ 2.9M (9.5 ft)
Sample pump and D.O. probe elevation @ 4.2M (14.0 ft)

**Aeration System Details**

A coarse bubble diffuser system consisting of 205 stations [two 13MM (1/2 inch) diameter holes]. The stations are spaced 1.2M (4 ft) center to center on laterals which are on 1.4M (4.5 ft) centers.

Figure 1. Test setup for the radiotracer and off-gas oxygen transfer tests.

TABLE 1. OPERATING CONDITIONS FOR RADIOTRACER AND OFF-GAS OXYGEN TRANSFER TESTS

| Test Parameter | Level |
|---|---|
| Aeration volume | $2.5 \times 10^6$ ℓ (0.66MG) |
| Wastewater flow | 22 ℓ/sec (350 gpm) |
| Recycle flow | 25 ℓ/sec (400 gm) |
| Aeration stations | 205 (7.1 ℓ/sec/sta) |
| Air flow rate | 1450 ℓ/sec (3075 scfm) |
| Water temperature | 21.8°C |
| MLSS | 6100 mg/ℓ |
| TDS | 5000 mg/ℓ ($\beta$ = 0.94) |
| $O_2$ uptake | 198 mg/ℓ/hr |
| D.O. | 0.0 mg/ℓ |

## Radio Tracer Results

Each sample from the four sample locations were analyzed in the laboratory for tritium and kryton-85. Typical counting efficiencies for this study were about 25 percent for tritium and about 82 percent for krypton-85.

As previously discussed, this study deviated from the normal tracer measurement procedure of having both the tritium and krypton-85 gas released from a single homogeneous bottle. Due to this deviation, the basic assumption of identical concurrent mixing for both tracers is invalidated and only krypton:tritium ratios measured after complete mixing was achieved were useful for calculating the krypton-85 transfer rate. This complete mix criterion gives approximately thirty minutes of useful gas transfer data for Stations 1, 2 and 3 and about fifteen minutes of useful data for Station 4. A least squares analysis of the logs of useable krypton:tritium concentration ratio data vs time was performed for each sampling station. The log-slope from each plot is the krypton transfer rate coefficient, $\alpha K_{kr}$, which is converted to the equivalent $\alpha K_{La}$ value by the relationship $(K_{kr} : K_{La}) = 0.83$ (14).

The computed oxygen transfer rate coefficients, along with corresponding statistical fit parameters are shown in Table 2. The average field oxygen transfer rate, $\alpha K_{laT}$, was determined to be 4.85 $Hrs^{-1}$ with a range from 4.72 to 5.00 $Hrs^{-1}$ for the four sample locations.

## Off-Gas Analysis Results

The hood data from the twelve aeration tank locations were analyzed to determine the percent oxygen transfer at each location. These results were then averaged, based upon the weight of the unit gas rate ($\ell/sec/M^2$) to obtain the overall average tank transfer efficiency of 7.7 percent. Since the extrapolated air flow collected by the off-gas hood was slightly higher than that measured going into the aeration tank, a range of transfer rates was calculated. The air collected by the hood averaged 3.8 $\ell/sec/M^2$ vs 3.5 $\ell/sec/M^2$ measured (by an orifice plate) coming into the tank. Using this air flow range and the overall transfer efficiency determined for the tank (7.5 percent), the $\alpha K_{laT}$ level was determined using Equation 7 below:

$$\alpha K_{LaT} = \frac{(OTE)(C.F.)(Q_A)}{(\beta C^*_{\infty T} - C_R)D} \tag{7}$$

where:

OTE = Oxygen Transfer Efficiency (Decimal)

C.F. = Conversion Factor = $1.0 \times 10^3$

TABLE 2. OXYGEN TRANSFER TEST RESULTS FOR THE RADIOTRACER METHOD

| Station[a] | $\alpha K_L$[a] @ T = 21.8°C | Coefficient of Variation, $r^2$ |
|---|---|---|
| 1 | 4.86 $hrs^{-1}$ | 0.997 |
| 2 | 5.00 $hrs^{-1}$ | 0.989 |
| 3 | 4.72 $hrs^{-1}$ | 0.987 |
| 4 | 4.81 $hrs^{-1}$ | 0.950 |
| Average | 4.85 hrs-1 | 0.981 |
| Standard Deviation | 0.12 $hrs^{-1}$ | 0.021 |

[a] Note: Stations are the same as sample pump locations.

$Q_A$ = Airflow Rate Per Unit Area ($\ell$/sec/$M^2$)

$\beta C^*_{\infty T}$ = Field DO Saturation Concentration (mg/$\ell$)
(Note: $C^*_{\infty T}$ = 10.0 mg/$\ell$ determined in previous clean water tests; $\beta$ = 0.94 from Table 1.)

$C_R$ = Residual DO Concentration = 0 mg/$\ell$

D = Depth = 6.1M

Table 3 shows the range of field oxygen transfer rates, $\alpha K_{LaT}$, to be 4.75-5.09 $Hrs^{-1}$ for the off-gas analysis. Comparing this range with that from the radiotracer test (4.72-5.00 $Hrs^{-1}$), there is excellent agreement between these two test methods. Thus, for this submerged aeration application both of these testing methods appear to be equally appropriate.

## CONCLUSIONS

Aeration testing with respiring activated sludge systems is not easily carried out and is not recommended over clean water aeration testing for verifying aeration performance specifications. However, with careful data collection and evaluation, it is possible to obtain reasonably valid results under process conditions. After review of current testing experience and the inherent errors of each test, the following conclusions are appropriate:

- Batch endogenous testing procedures are more accurate than continuous testing techniques for estimating field transfer rates of various aeration equipment.

- The main limitation of continuous testing is the inability to accurately measure the biological oxygen uptake rate.

- For many applications, specifically for surface aeration installations, the DO concentration may vary spatially within the tank at a given time during testing under continuous loading conditions; therefore, conventional testing approaches may not be valid for these applications. Further study is required for such installations.

- For surface aeration equipment, steady state batch endogenous testing is preferable to non-steady state batch endogenous testing.

- For submerged aeration equipment, non-steady state batch endogenous testing is preferable to steady state batch endogenous testing.

- Batch endogenous desorption testing using hydrogen peroxide has been demonstrated as an effective technique for measuring field oxygen transfer in surface and submerged aeration systems.

- The mass balance approach for determining field oxygen transfer rates requires extensive data collection. For activated sludge systems, the approach appears to have value in small, well-mixed systems but is simply

**TABLE 3. OXYGEN TRANSFER TEST RESULTS FOR THE OFF-GAS ANALYSIS TECHNIQUE**

| Data Description | Mean[a] OTE | Unit Air Flowrate | $\alpha K_L a_T$[b] |
|---|---|---|---|
| Hood Collected Air Flow Rate | 7.7% | 3.8 l/sec/M$^2$ | 5.09 hrs$^{-1}$ |
| Measured Influent Air Flow Rate | 7.7% | 3.5 l/sec/M$^2$ | 4.75 hrs$^{-1}$ |

[a] Mean OTE: Average oxygen transfer efficiency (%) under actual field operating conditions (see Table 1).

[b] $\alpha K_L a_T$: Average oxygen transfer rate parameter (hrs$^{-1}$) under actual field operating conditions (see Table 1).

not as accurate as other tests for most systems. For ASB systems, this approach may be the only practical method available for testing. Therefore, with reasonable care in data collection and analysis, the mass balance approach can be useful for evaluating aeration equipment performance.

- The off-gas analysis method for field oxygen transfer measurement has had limited application to date. One current application of this technique is with covered tank, pure oxygen activated sludge systems. However, in view of advancing technology in oxygen concentration measurement equipment, and the progressive increase in efficiency of aeration systems generally, this method deserves further evaluation. Preliminary findings using this approach indicates high potential for this method in bottom coverage diffused aeration applications.

- The tracer method for field oxygen transfer determination remains relatively untested. Preliminary findings using this technique are encouraging. The approach requires and deserves further evaluation.

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## DISCLAIMER

**The work summarized herein was previously reviewed and approved for publication by the U.S. Environmental Protection Agency (EPA-600/2-83-002).**

# AERATOR PERFORMANCE EVALUATION - THE ONTARIO PERSPECTIVE

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## ABSTRACT

Aeration equipment employed in the activated sludge process is usually the single largest consumer of energy in a wastewater treatment plant. Consequently, the selection of efficient aerators is an essential aspect of treatment plant design. Although oxygen transfer efficiency is only one of several major selection criteria, it has a foremost impact on the long-term energy requirements for plant operations.

This paper discusses acceptance and performance testing procedures adopted by the Ontario Ministry of the Environment in the specification and selection of aeration equipment for activated sludge wastewater treatment facilities in Ontario.

## INTRODUCTION

The requirement of effective aeration devices to maintain an aerobic environment in the activated sludge process is universally recognized. Such an aerobic environment not only implies a sufficient dissolved oxygen concentration, but also adequate contact between wastewater, microorganisms and the dissolved oxygen. Usually, the aeration device must also provide for this contact.

Although there are many factors which enter into the selection of the aeration device for a particular project, the primary measure of effectiveness of an aeration device is its energy efficiency.

As with other energy intensive industries, energy conserving designs and operations are receiving increased emphasis in the wastewater treatment field. Aeration equipment employed in the activated sludge process is usually the single largest consumer of energy in a wastewater treatment plant, normally accounting for 60-80 percent of the total power demand. Effective aeration, therefore, implies maximum oxygen input with a minimum horsepower requirement. Oxygen input is measured in terms of oxygen transfer efficiency.

Few, if any, unit operations in the wastewater treatment field receive the design effort, intensity of performance testing or general operational attention reserved for aeration devices. The great variability in aeration type and design, and the large number of manufacturers, ensures a keen competition.

Thus, aeration testing methods and results are scrutinized and critically reviewed to an uncommonly high degree by all concerned: the manufacturer, the design engineer, the prospective owner and in many cases, the regulatory agencies as well as the competition.

## ACCEPTANCE OF AERATION DEVICES FOR USE IN ONTARIO

Most aeration devices employed in municipal wastewater treatment facilities in Ontario receive prior Ministry of the Environment (MOE) acceptance. Such acceptance, although not essential for marketing purposes, is generally looked upon by consulting engineers in Ontario, and indeed across Canada, as an informal accreditation of the equipment.

Therefore, MOE acceptance is generally sought by the manufacturer/supplier when a new aeration device is presented for consideration in the Canadian market. In such cases, manufacturer's claims of the equipment are initially reviewed. If the claims are considered reasonable and in line with similar known equipment, general acceptance for use in Ontario is given. Such general acceptance is not to be construed as final approval as each project must receive MOE approval on its own merits for the particular conditions applying.

If the manufacturer's claims are questionable or exceed generally accepted standards, or if the aeration device works on an unfamiliar principle, performance testing may be requested. Such testing may be conducted outside the province, provided that an accepted testing procedure is followed and adequately documented, or the testing can be conducted under MOE supervision at the Ontario Experimental Facility operated by the Wastewater Treatment Section, MOE. The latter situation not only provides for MOE certification of the test results, but also provides an opportunity for the manufacturer/supplier to introduce both himself and his equipment to local consultants and MOE personnel.

Should the MOE reject the acceptance of the device, it may still be marketed in Ontario for private and municipal projects. However, as mentioned above, in general, consultants in Ontario and elsewhere in Canada look to MOE acceptance as a prerequisite in aeration device selection.

## SELECTION OF AERATION DEVICES IN ONTARIO

Generally, the selection of aeration equipment in Ontario is the responsibility of the design consulting firm for the project. To the consulting engineer, the selection of aeration equipment involves the development of a total aeration system capable of serving a variety of needs within the wastewater treatment plant. Other air requirements such as for pre-aeration, aerated grit removal, channel aeration, dissolved air flotation, etc., must be considered, and the entire biological treatment system design will definitely influence aeration equipment type selection. The most efficient and economical transfer of dissolved oxygen is only one of several major selection criteria. Some of the other criteria specific to the aeration device include capital and installation cost, aeration capacity, equipment reliability and mixing performance. Temperature loss or addition due to the aeration may also be a factor in northern climates as well as aerosol minimization where facilities are covered.

The aerator is usually the only source of mixing within the aeration tanks. It must provide sufficient velocity within the tanks to maintain all biological floc in suspension and thorough mixing to provide a uniform and rapid contact between the wastewater, the introduced air (oxygen) and the microorganisms. Such mixing must at the same time, keep biological shear to a practical minimum.

In some MOE managed projects, the aeration equipment may be preselected. Preselection may be requested for a variety of reasons including cost-savings, time-saving, assurance of quality equipment and possibility of pre-installation performance testing.

The MOE has been involved in equipment selection for MOE managed projects for many years. As the owner and operating authority of a facility, it is felt essential to select quality equipment from reliable suppliers in order to minimize subsequent operating problems. While the consulting engineer's recommendation is largely relied upon, the MOE reserves the right to make the final decision in the selection of major pieces of equipment.

The MOE maintains an information list of prequalified, acceptable suppliers and manufacturers of major equipment including aeration devices. As only prequalified suppliers are invited to submit quotations for an item of equipment that is to be preselected, other suppliers and manufacturers are continuously seeking general acceptance by the Ministry of various equipments not yet ruled as acceptable for MOE capital works. Consequently, procedures such as the oxygen transfer testing procedure have been established to provide uniformity in evaluating classes of equipment.

In cases of preselection, the consulting engineer prepares the necessary specifications for MOE review. The consultant then invites quotations from prequalified suppliers, evaluates the quotes and makes appropriate recommendations to the MOE, who makes the final decision on selection.

The equipment may then be pre-ordered by purchase order directly by the MOE if long deliveries or price increases are anticipated, arranged for through a committing letter of intent to permit manufacturing to begin, or the preselected equipment may be ordered by the general contractor following successful tendering.

Through the preselection process, certain major pieces of equipment are selected in such a way that public funds are spent in a most responsible manner.

## PERFORMANCE EVALUATION AND COMPLIANCE TESTING

Because of the many and varied types of aeration devices on the market today, some means of comparing their relative efficiencies is essential.

Although oxygen transfer efficiency is only one of the several major criteria used in the selection of an aeration device, it is extremely important in assessing long-term energy requirements.

A standard oxygen transfer testing procedure has been developed and adopted by the Ontario Ministry of the Environment for use in comparing the oxygen transfer efficiencies of different aeration devices. The testing procedure was developed and refined over a period of almost 20 years of testing by the Ontario Water Resources Commission and now Ontario Ministry of the Environment. It has seen many modifications and refinements but is now considered to reflect the current state-of-the-art.

This oxygen transfer test is really a clean water test carried out under standard, but rather ideal conditions and can be used only in determining the oxygen transfer efficiency relative to other aeration devices. Compliance testing may, and indeed frequently is specified by the consulting engineer to determine whether or not design specifications are met under field conditions. Compliance testing is conducted following installation of the aeration device and may be used to determine modification requirements in a multi-unit facility. This is one instance in which equipment preselection can be extremely beneficial. Compliance testing procedures are site specific and are generally developed to meet the needs of a particular project.

## MOE OXYGEN TRANSFER TESTING PROCEDURES

While on the surface it may seem a simple enough matter to devise a standard oxygen transfer testing procedure, in practice the difficulties are diverse due to the many variables which must be considered. There are currently over two dozen oxygen transfer test methods in use including those of the MOE, ASCE, ASTM, EPA and of individual consulting firms. Although many of the methods are very similar and vary only in detail, such detail, particularly in the method of calculation, can influence final results to such a degree that judicious selection of method could enable one to obtain predestined transfer results. From past MOE experience, many manufacturers' performance claims are exaggerated when compared to MOE field tests on the same aeration device. This is usually a result of the claimed performance being based upon a method which maximizes apparent performance.

The Ministry of the Environment oxygen transfer acceptance testing procedure is based on the fullscale, non-steady state tapwater test. Although the test procedure is presented in Appendix A as "Procedure for Evaluating Aerator Performance", general comments concerning the test are addressed below.

### Site Preparation

Testing is conducted at the Ontario Experimental Facility in a large scale concrete tank which is described in Appendix B.

The test tank has length, width and depth measurements of 27 ft x 24 ft x 14 ft with a capacity of about 9,000 $ft^3$. Small pilot-scale tests are not acceptable in Ontario because of possible side wall effects and scale-up problems. The aeration device is installed in accordance with the manufacturer's instructions and usually run under operating conditions (submergence, air flow rate, etc.) selected by the manufacturer to maximize performance in that particular test tank. A range of operating conditions may be used.

## Test Medium

Municipal tapwater is used as the medium in aerator performance testing. Although water chemistry can be an important factor in oxygen transfer, and such materials as iron salts and other dissolved solids, even in relatively small amounts can enhance or depress oxygen transfer performance, municipal tapwater supplies are generally of relatively high and consistent quality. For tapwaters in Ontario, ∝ and β are considered to be unity, however, the test engineer should be satisfied from past knowledge, or test water analyses, that no impurities are present to interfere with the tests.

Although some test procedures call for a standard test water temperature of 20°C, this is not always practical to achieve. Acceptance tests have been carried out in Ontario at water and ambient air temperatures as low as 1°C and -30°C, respectively. As long as the temperature correction to standard conditions is judiciously applied, the MOE testing procedure applies to any reasonable test water temperature.

## Deoxygenation

In the MOE testing procedure, the aeration equipment is run at a steady-state for at least 30 minutes prior to testing to allow establishment of the hydraulic regime, constant temperature and pressure of air supply, steady-state power draw, etc.. During this period, Cobalt as cobalt chloride is added as a catalyst at a concentration of 0.25-0.5 mg/L. Sodium sulfite is then added, preferrably as a solution but permissibly as a slurry (cold weather conditions), to the test tank as rapidly as possible using a submersible pump and spraying over the entire tank surface. Sulfite additions are made in excess of stoichiometric requirements, the magnitude of which depends on expected oxygenation and mixing rates.

## Dissolved Oxygen Measurement

Reoxygenation of the dissolved oxygen (DO) depleted tapwater is measured by means of several DO meters whose probes are located at selected points in the tank to check for any significant variation in the oxygen transfer coefficient.

Several other testing procedures recommend or require the use of the Winkler DO test requiring withdrawal of discrete samples or the use of submersible pumps. However, after exhaustive testing of most of the available DO measurement methods over a twenty-year period, the MOE has full confidence in DO meters provided they are in good condition, accurately calibrated and show fast response times. The DO meters used in the MOE procedure, usually 4 or 5 with a minimum of four, must be accurately calibrated each day before testing begins.

During the test run, the tank DO is allowed to reach 90% of the saturation value at the test temperature before sulfite is added for the next run.

## Dissolved Salt Limitation

As the addition of sulfite increases the dissolved salts concentration in the tapwater, incremental addition of sulfite is allowed to a maximum of 1,500 mg/L total dissolved solids before the testwater must be changed.

## POWER MEASUREMENT AND EXPRESSION OF RESULTS

During the test runs, power and/or air flow readings are taken to measure watts, volts, amps, power factor and scfm supplied. Calculation of the test results is by the log deficit method as outlined in Appendix 1.

Performance is described in terms of capacity (lb $O_2$/h), electrical efficiency (lb $O_2$/kWh) and % efficiency. It should be emphasized that power factor is measured in these tests. It has been found that the power factor is frequently as low as 0.5 rather than the normally assumed 0.85 value, considerably affecting power draw calculations.

## Mixing

Mixing performance, although a very important consideration in aerator selection, is a very difficult parameter to define. Fortunately, it usually follows that should oxygenation performance be good, then mixing will also be satisfactory. Standard requirements for mixing are that a velocity of 0.25 m/s is maintained at the tank bottom to suspend solid material. In some instances where such velocities are achieved, solids settling reportedly still occurs. Lately, several aeration systems have been marketed which provide finally diffused air through a grid of diffusers covering the total tank bottom. Such systems do not provide a high unidirectional bottom velocity but do provide adequate vertical shear to prevent bottom deposition. Thus, although velocity measurements are a useful indication of mixing performance, they are not always applicable. As an indicator of mixing throughout the whole test tank, the Ministry of the Environment standard requires that the measured $K_La$ for each sampling point shall not vary more than ± 10% from the mean $K_La$ for all sampling points. This requirement is fairly lenient and most tests conducted to date have fallen well within these limits; some mixing devices have, however, failed to meet the mixing requirements in MOE acceptance testing and were, therefore, deemed unacceptable. Other devices have only failed to meet the mixing criteria under specific operating conditions, and the testing has defined equipment limitations.

## SUMMARY

In summary, proper selection of aeration equipment is essential to the successful and efficient operation of the activated sludge wastewater treatment process. The large number and diverse nature of aeration devices available on the market necessitate a standardized approach to aerator evaluation and selection.

Ontario has adopted such a standardized approach including performance evaluation and compliance testing involving a detailed oxygen transfer test procedure. While some criticism may be directed at the conservative nature of the MOE method, it must be realized that all aeration devices receiving MOE acceptance are evaluated and rated in a similar manner.

## APPENDIX A - PROCEDURE FOR EVALUATING AERATOR PERFORMANCE

### 1) Test Conditions and Initial Preparations

Prior to conducting oxygen transfer testwork, discussions shall be held between MOE personnel and the equipment supplier/consultant to ensure that the installation conforms to required specifications and that subsequent test results obtained will definitively reflect the performance of the aeration device under the test conditions.

All aerator acceptance tests will be conducted on clean tapwater; if requested, information on the water supply chemistry shall be provided by the MOE to the supplier/consultant.

On the basis of all available data, the supplier/consultant will provide the MOE with a performance claim or guarantee of transfer capacity in kg $O_2$/h and kg $O_2$/kWh, and show the test conditions of submergence, aerator speed, etc. relating to the guarantee.

After aerator installation, the test tank shall be thoroughly cleaned and filled with fresh tapwater. From visual and/or chemical inspections, all parties shall agree upon the suitability of the testwater for aerator performance characterization.

### 2) Deoxygenation of Testwater

Cobalt free anhydrous sodium sulphite, $Na_2SO_3$, shall be used to deoxygenate the testwater. The quantity of $Na_2SO_3$ added for each run will be based upon the equation:

$$Na_2SO_3 + 1/2\ O_2 \longrightarrow Na_2SO_4$$

which shows that 7.9 kg of $Na_2SO_3$ will remove 1 kg of dissolved oxygen from the testwater. From an initial dissolved oxygen measurement on the testwater, the weight of oxygen present will be calculated and the stoichiometric quantity of $Na_2SO_3$ plus an adequate excess added to achieve zero DO. The excess $Na_2SO_3$ added will be based on the expected oxygen transfer rate of the aerator.

The $Na_2SO_3$ will be dissolved in an external tank mixed by a propeller or recirculating pump, and will be introduced to the testwater as rapidly as possible by a submersible pump spraying the chemical equally over the entire surface of the test tank.

While not preferred, it is permissible in some instances to add the $Na_2SO_3$ to the test tank as a slurry.

3) Cobalt Catalyst

Cobalt shall be used to catalyze the deoxygenation reaction. The cobalt concentration used shall be at minimum 0.25 mg/L and at maximum 0.5 mg/L.

Cobalt chloride, $CoCl_2 \cdot 6H_2O$, shall normally be used as the source of cobalt catalyst. The cobalt salt will be completely dissolved in tapwater and added to the testwater during steady-state aerator operation at least 30 minutes prior to the first run.

4) Physical Parameters

The testwater volume shall be maintained constant during each test series. Where several water levels are to be tested, the greatest shall be conducted first to allow assessment of motor/blower suitability under high load conditions.

The aeration device shall be operated at test conditions for at least 30 minutes prior to starting any test run to allow establishment of the hydraulic mixing regime, steady power draw on drive motors, etc..

Under no circumstance will the aeration device be stopped from commencement of the above 30 minute stabilization period until the end of the test run. A maximum of 10 runs, or concentration of 1500 mg/L $Na_2SO_4$, whichever occurs first, shall be allowed on any batch of testwater.

5) Sampling and DO Measurement

A minimum of four DO measuring stations shall be selected by discussion between all parties. All efforts will be made to select measuring stations which best represent the overall tank characteristics.

The primary DO measurement method will be by fast response membrane type probes. These shall be capable of >90% needle deflection from DO saturation towards zero DO in 45 seconds or less. All meters will be accurately calibrated daily against the Winkler DO method as outlined in Standard Methods. While it is permissible to use DO recorders, it is preferred that test calculations be based on readings taken directly from the meters.

Although not required, measurement of oxygen transfer performance by removal of discrete samples for Winkler analyses is acceptable.

At least 10 DO readings will be made during each test at each measurement station. The maximum allowable time between each reading will be 1 minute.

During each run, DO readings will be continued to at least 90% of the saturation value.

## 6) Data Interpretation

A minimum of three runs at each test condition will be made to define performance at that condition. At least three consecutive runs at the same conditions shall be within ±10% of the average of the runs. Failure to comply will necessitate further runs at the test condition to define the performance obtained.

For each run, $K_La$ values will be calculated for each DO measuring station. The $K_La$ value for each station shall not vary more than ±10% from the mean value for each test. Greater variation will indicate incomplete mixing and invalidate the test.

The raw data will be analyzed using the log deficit method. DO values less than 20% and greater than 90% of the test DO saturation value may be discarded in the calculations. It is preferred that data will be depicted graphically in addition to, or opposed to computer program analyses.

### Calculation of $K_La$ for Mechanical Surface Aerators

In the case of mechanical surface aerators, the overall coefficient of oxygen transfer, $K_La_{20}$, shall be calculated by:

$$K_La_{20} = \frac{1}{(t_2 - t_1)} \ln \frac{(C_s - C_1)}{(C_s - C_2)} \theta^{20-T}$$

where: $K_La_{20}$ = overall coefficient of oxygen transfer at 20°C

$C_s$ = Standard Methods value for dissolved oxygen saturation at test temperature, T°C

$C_1, C_2$ = DO concentration at times, $t_1$ and $t_2$

$\theta$ = temperature correction coefficient = 1.024

$T$ = temperature of testwater, °C

### Calculation of $K_La$ for Submerged Aeration Devices

In the case of submerged aeration devices, the overall coefficient of oxygen transfer, $K_La_{20}$, shall be calculated by:

$$K_La_{20} = \frac{1}{(t_2 - t_1)} \ln \frac{(C_{sm} - C_1)}{(C_{sm} - C_2)} \theta^{20-T}$$

where $C_{sm}$ is the mid-depth dissolved oxygen saturation saturation value and is calculated by:

$$C_{sm} = C_s \frac{[P + 0.5x] - p}{760 - p}$$

where: $C_s$ = the Standard Methods DO saturation value at test temperature, T°C

$p$ = saturated water vapour pressure at test temperature, T

$P$ = barometric pressure, mm

$x$ = hydrostatic head above air release point

$\theta$ = temperature correction coefficient = 1.024

For each test, $K_La_{20}$ values for all DO measuring positions shall be averaged and used to calculate oxygenation performance.

In the case of mechanical surface aerators, performance shall be expressed in terms of transfer capacity, kg $O_2$/h, and electrical efficiency, kg $O_2$/kWh.

In the case of submerged aeration devices, performance shall be shown in terms of capacity, kg $O_2$/h, % efficiency, and also as kg $O_2$/kWh, where kW is blower motor wire power.

The calculated mid-depth oxygen saturation value shown previously will be used to calculate the above.

$\alpha$ and $\beta$ factors shall be assumed as unity in tapwater tests.

7) Power Measurement

In the case of mechanical surface aerators, power shall be measured by an indicating polyphase wattmeter. Manufacturer's calibration data shall be used to determine the motor and reducer efficiencies. In addition, the voltage in all three phases shall be shown, the current in at least two phases, and the power factor determined. Instrument accuracy shall be 1% or better, current transformers 0.5% or better.

A similar instrument shall be used to measure blower power draw in the case of submerged devices. In addition, air flow rates should be accurately measured using an in-line orifice plate or other acceptable means.

All power and air flow readings shall be taken when the aeration device is at steady-state operating conditions. Several sets of readings shall be taken during each oxygen transfer test.

8) Mixing Performance

Compliance with the ±10% $K_La$ variation shown in Item 6, is acceptable as proof of adequate mixing performance. However, in cases where actual velocity measurements are made, these shall be conducted using an Ottmeter or by a Gurley meter.

Minimum liquid velocity at all points in the tank shall be 0.25 m/s at full motor load.

## APPENDIX B - OXYGEN TRANSFER TEST TANK DESCRIPTION - ONTARIO EXPERIMENTAL FACILITY

**The test cell is an isolated bay within an existing 100 ft (≈30 m) diffused aeration tank of the Ontario Experimental Facility, a 5.0 MIGD activated sludge plant maintained by the MOE for research and training purposes. The bottom longitudinal corners are filleted and all construction is concrete including the isolation wall installed to form the test cell. A Y-wall separating the test cell from a parallel common wall tank limits effective test depth to 14' (4.3 m).**

**Test tank specifications as as follows:**

| | |
|---|---|
| **Dimensions:** | **L 27 ft (8.2 m); W 24 ft (7.3 m); D 14 ft (4.3 m)** |
| **Capacity:** | **9,072 ft$^3$ (257 m$^3$)**<br>**56,609 IG (67,931 USG)** |
| **Air Supply:** | **Blowers - 2 Sutorbilt 30 HP, 980 cfm @ 7.5 psi each;**<br>**- 1 Cycloblower 50 HP, 1940 cfm @ 18 psi.** |
| **Electrical:** | **575 V and 110 V outlets on-site (stepdown transformer can be installed for 440 V).** |
| **Water:** | **Tapwater available from 5 nearby hydrants.** |
| **Laboratory:** | **Both permanent and mobile laboratories available on-site.** |
| **Hardware & Analytical Equipment:** | **Winkler DO chemical and bottles supplied;**<br>**Stream Sampler supplied;**<br>**DO meters available (YSI and EIL);**<br>**Sulfite makeup tank complete with mixer and pump supplied;**<br>**Electrical test measurement kit available;**<br>**Truck mounted hoist available.** |

### DISCLAIMER

**The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency, and no official endorsement should be inferred.**

# COMPARISON OF DUAL NONSTEADY STATE AND STEADY STATE TESTING OF FINE BUBBLE AERATORS AT WHITTIER NARROWS PLANT, LOS ANGELES

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## ABSTRACT

The dual nonsteady state technique has been developed to evaluate aeration system oxygen transfer coefficients and saturation values under process conditions. This paper provides the results of application of the technique to a long narrow aeration tank at the Whittier Narrows plant, Los Angeles, Ca.

## INTRODUCTION

During the summer of 1981 aerator testing using different techniques was conducted at the Whittier Narrows plant, Los Angeles County. The testing program was a cooperative effort between the ASCE Committee on Oxygen Transfer Standards and the Los Angeles County Sanitation District (LACSD) which was conducting an EPA supported study on the efficiency of various aeration devices at the plant. Three techniques were used for testing during this period; (1) the dual nonsteady state analysis, (2) the steady state analysis and (3) the off-gas analysis.

This report presents the results of the first two testing techniques, the dual nonsteady state and the steady state analyses. The main body of the report summarizes the study procedures and results of the two techniques.

## DESCRIPTION OF TESTS

The Whittier Narrows wastewater treatment plant is a facility containing primary clarifiers, long narrow aeration tanks, secondary clarifiers, chlorination tanks and tertiary filters. It treats mainly domestic sewage at a relatively constant flow rate over the day. The influent COD concentration exhibits a marked diurnal variability with minimum values occurring in the late morning hours, increasing during the day to a maximum value which remains relatively constant from the early afternoon hours through the evening hours (Figure 1). To attain a relatively constant $O_2$ uptake rate during the tests, non steady state data was obtained in the evening starting at the earliest around 5:30 P.M. with some of the tests completed after midnight.

Figure 2 shows a schematic of the aeration tank layout at Whittier Narrows. Fine bubble diffusers were located in the first two bays, one with full floor coverage discs and the other with double spiral rolls tubes, both constructed in a tapered aeration mode with approximately 45% of the

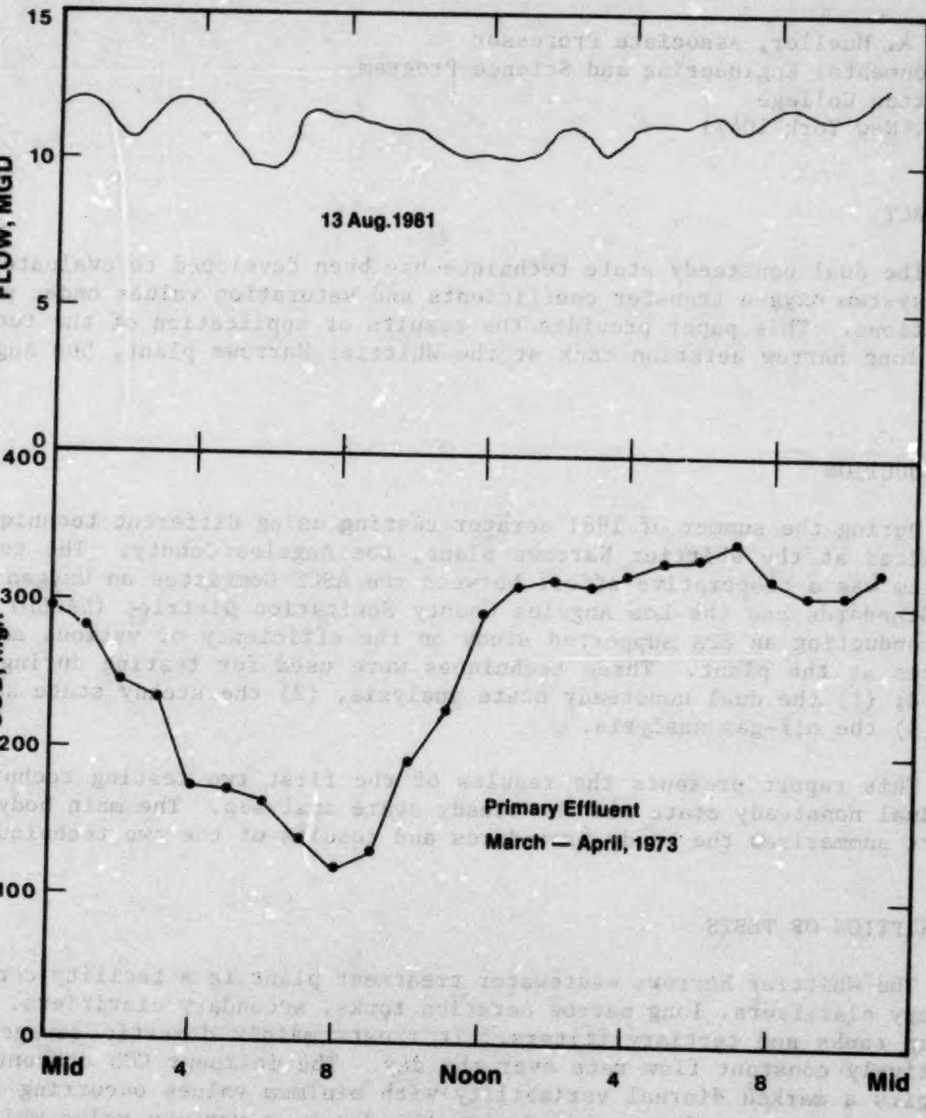
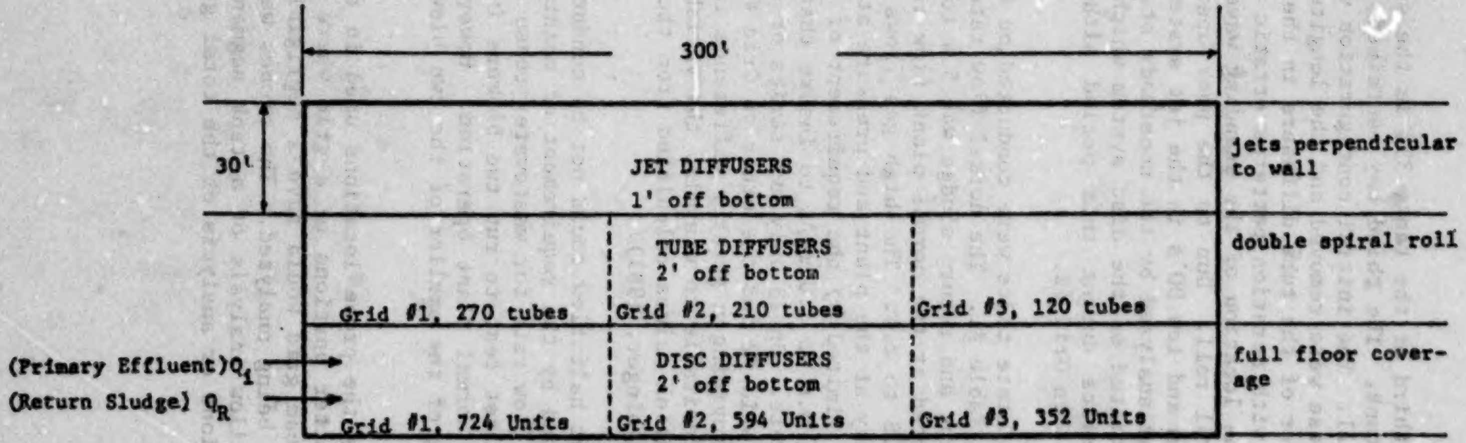


**Figure 1.** Diurnal Variability of Wastewater Flow and COD Concentrations for Whittier-Narrows Plant

Figure 2, Whittier Narrows Plant - Aeration Tank Layout


300'
30'
JET DIFFUSERS
1' off bottom
jets perpendicular to wall
TUBE DIFFUSERS
2' off bottom
Grid #1, 270 tubes
Grid #2, 210 tubes
Grid #3, 120 tubes
double spiral roll
(Primary Effluent) $Q_1$
(Return Sludge) $Q_R$
DISC DIFFUSERS
2' off bottom
Grid #1, 724 Units
Grid #2, 594 Units
Grid #3, 352 Units
full floor coverage

Water Depth for Disc Diffusers = 14.3-14.4' diffusers in the first third of the tank, 35% in the second third and 20% in the last third of the tank. The third bay contained jet diffusers located perpendicular to the wall. The initial configuration was jet clusters in the center of the tank. These were removed and the longitudinal header installed in their place. A number of the tube diffusers in the second bay had broken off prior to the study with aeration patterns erratic down the length of the tank as evidenced by the location of the standing wave on the tank surface due to the double spiral roll. Due to the poor operation of the tube diffuser aeration system and low DO's in the jet system during the summer of 1981, neither system was analyzed by the nonsteady state testing technique. All testing was concentrated on the disc system which exhibited excellent operational characteristics during this period although unequal flow distribution was evident in Grid #2.

Six dual nonsteady state tests were conducted on Grid #3 and one test on Grid #2 as indicated in Table #1. The normal flow rate to the system including both primary effluent and return sludge was 5.4 to 6.0 MGD. Five of the seven tests were conducted at the normal plant flow rate allowing gas flow ratios (high/low) of 1.8 to 2.4. The high gas flows were limited by the available blower capacity at the plant and pressure at the blower discharge. The low gas flows were limited by the requirement of maintaining a minimum D.O. concentration of ~ 0.6 to 1.0 mg/ℓ to insure that oxygen concentration does not limit uptake rate. On 23 July two series of tests were conducted at half flow rate, one on Grid #2 and the other on Grid #3. This allowed lower gas flows to be used providing a greater difference in the high to low gas flow ratios. This is desirable to minimize the variability in estimated $O_2$ uptake rates and saturation values calculated from the dual non steady state analysis (Mueller and Rysinger, 1981).

Additional tests at half flow could not be conducted due to the lack of plant flexibility imposed by the requirement of maintaining a high quality effluent at the normal flow rate for wastewater reuse downstream. An attempt was made prior to the last test to run two blowers in parallel instead of just the one used for normal plant operation. However this resulted in overheating and surging of the smaller of the two blowers which then had to be shut down.

Figure 3 indicates the probe locations used in the tests. They were concentrated in the latter portions of a grid where the denser diffuser placement of the upstream grid would have a negligible effect on transfer efficiency of the grid being analyzed. The probes were normally spread at regular intervals to allow analysis of a tank segment knowing inlet and outlet boundary conditions or analysis of the total grid segment assuming a plug flow mode.

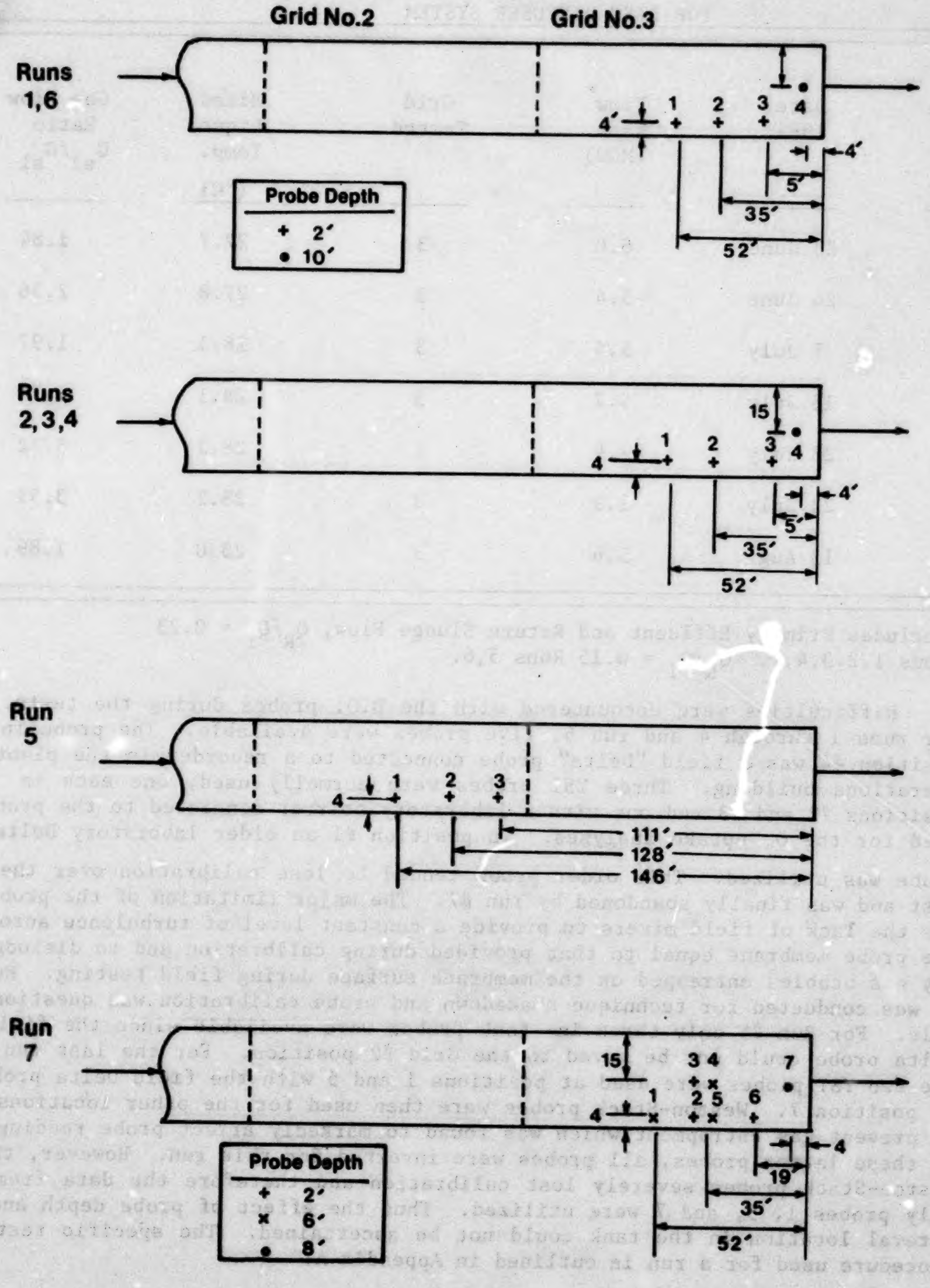
## TABLE 1. DUAL NONSTEADY STATE TEST CONDITIONS FOR DISC DIFFUSER SYSTEM

| Run | Date 1981 | Flow* Rate (MGD) | Grid Tested | Mixed Liquor Temp. (°C) | Gas Flow Ratio $G_{s2}/G_{s1}$ |
|---|---|---|---|---|---|
| 1 | 23 June | 6.0 | 3 | 27.7 | 1.84 |
| 2 | 24 June | 5.4 | 3 | 27.8 | 2.36 |
| 3 | 7 July | 5.4 | 3 | 28.1 | 1.97 |
| 4 | 13 July | 5.7 | 3 | 28.1 | 1.78 |
| 5 | 23 July | 2.9 | 2 | 28.1 | 3.72 |
| 6 | 23 July | 2.9 | 3 | 28.2 | 3.51 |
| 7 | 13 Aug | 5.6 | 3 | 28.0 | 1.89 |

*Includes Primary Effluent and Return Sludge Flow, $Q_R/Q_i = 0.23$ Runs 1,2,3,4,7. $Q_R/Q_i = 0.15$ Runs 5,6.

Difficulties were encountered with the D.O. probes during the tests. For runs 1 through 4 and run 6, five probes were available. The probe in position #4 was a field "Delta" probe connected to a recorder in the plant operations building. Three YSI probes were normally used, one each in positions #2 and #3 and one with a laboratory stirrer connected to the probe used for the $O_2$ uptake analyses. In position #1 an older laboratory Delta probe was utilized. This older probe tended to lose calibration over the test and was finally abandoned by run #7. The major limitation of the probes was the lack of field mixers to provide a constant level of turbulence across the probe membrane equal to that provided during calibration and to dislodge any gas bubbles entrapped on the membrane surface during field testing. Run #1 was conducted for technique shakedown and probe calibration was questionable. For Run #5 only three in- tank probes were available since the field Delta probe could not be moved to the Grid #2 position. For the last run, the two YSI probes were used at positions 1 and 5 with the field Delta probe at position 7. Weston-Stack probes were then used for the other locations. To prevent gas entrapment which was found to markedly affect probe readings of these latter probes, all probes were inverted for this run. However, the Weston-Stack probes severely lost calibration and therefore the data from only probes 1, 5, and 7 were utilized. Thus the effect of probe depth and lateral location in the tank could not be ascertained. The specific test procedure used for a run is outlined in Appendix A.

Figure 3. D.O. Probe Locations in Disc Diffuser System


Grid No.2
Grid No.3
Runs 1,6
1
2
3
4
4'
4'
5'
35'
52'
Probe Depth
+ 2'
● 10'
Runs 2,3,4
15
1
2
3
4
4
4'
5'
35'
52'
Run 5
1
2
3
4
111'
128'
146'
Run 7
15
3 4
7
1
2 5
6
4
4
19'
35'
52'
Probe Depth
+ 2'
× 6'
● 8'

## DATA ANALYSIS TECHNIQUES

To date, the nonsteady state aeration testing techniques have been applied to either batch or completely mixed systems for both clean water and respiring systems testing. The equations required to analyze these systems are given in Appendix B for the dual non- steady state analysis. Neither one of these approaches applies to the long narrow aeration tank at Whittier-Narrows under process conditions. Initially it was anticipated that the tank could be divided into completely mixed segments with inlet and outlet conditions defined. If the change in concentration between the inlet and outlet boundaries was constant with time (constant $\Delta C$ analysis) during a non steady state test, then the analysis for $K_La$ reduced to the batch system. The latter runs showed that this was not the case, therefore requiring another technique for data analysis.

Upon completion of Run #7, a D.O. profile was taken down the length of the total aeration tank as shown in Figure 4. Significant $O_2$ concentration gradients existed throughout the tank especially in Grids 1 and 3, indicative of a plug flow mode rather than a completely mixed one.

To determine the best technique in which to analyze the nonsteady state data in a plug flow type system, an analytical solution of the equations describing the system were obtained. The following conditions were used for this solution based on the Whittier Narrows data:

(1) The initial tank concentrations were set at the steady state concentrations for the previous gas flow prior to changing gas flow for the nonsteady state profiles,

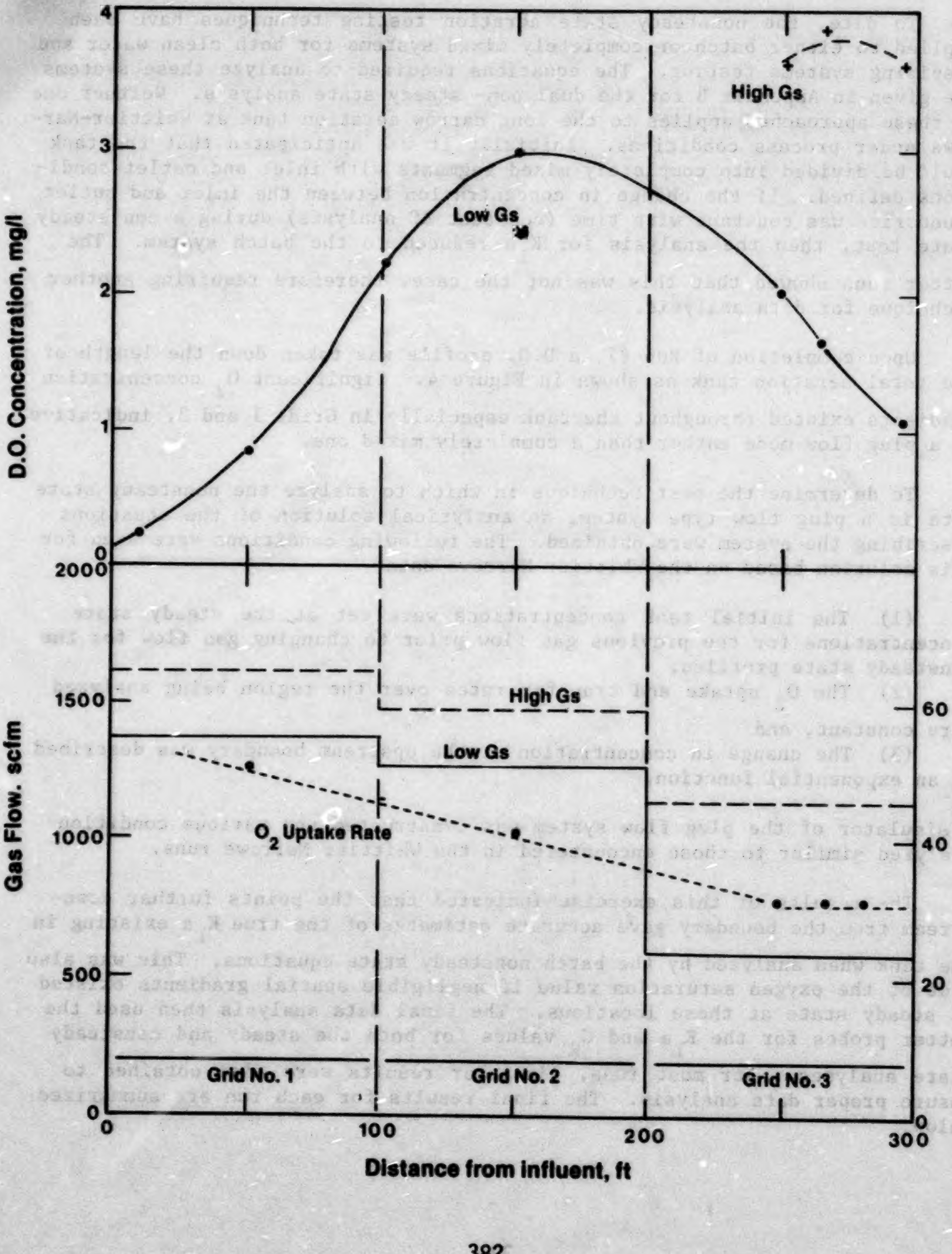
(2) The $O_2$ uptake and transfer rates over the region being analyzed were constant, and

(3) The change in concentration at the upstream boundary was described by an exponential function.

A simulator of the plug flow system was constructed and various conditions analyzed similar to those encountered in the Whittier Narrows runs.

The results of this exercise indicated that the points further downstream from the boundary gave accurate estimates of the true $K_La$ existing in the tank when analyzed by the batch nonsteady state equations. This was also true of the oxygen saturation value if negligible spatial gradients existed at steady state at these locations. The final data analysis then used the latter probes for the $K_La$ and $C_R$ values for both the steady and nonsteady state analyses. For most runs, simulator results were also obtained to insure proper data analysis. The final results for each run are summarized below.

Figure 4. Profiles of D.O. Concentration and Uptake Rate at End of Run #7 over Total Tank Length



## RESULTS

The gas flow measurements, average oxygen uptake rates and $O_2$ saturation values estimated from clean water data are presented in Table 2 for each run. Relatively low gas flows were obtained in Runs 5 and 6 due to the lower uptake rates occurring at half normal flow. The clean water saturation values were relatively constant since the water depth and temperature were similar for all runs. Two β factors, 0.95 and 0.99, were used for this data. The 0.95 value was common practice of the LACSD. However based on TDS values from 480 to 570 mg/ℓ, a β factor of 0.99 would be more applicable. Table 3 summarizes the dual nonsteady state results for each run. No $O_2$ uptake rates or saturation values could be obtained for Runs #2 and #3 since the low gas flow steady state oxygen value was less than zero and a reliable estimate of $K_La$ could not be attained. In Run #7 a spatial gradient still existed at the effluent end of the tank requiring an estimate of the steady state D.O. value for the uptake rate and saturation value determination. The steady state results are presented in Table 4. These were calculated using the saturation values estimated from clean water values in Table 2 with the steady state $O_2$ concentration in Table 3. For the high gas flows the higher β value of 0.99 yields 6 to 10% lower calculated $K_La$ values while the lower gas flows show a reduction of 3 to 4% due to the lower steady state D.O. concentrations and increased driving forces.

Figures 5 and 6 show the correlations of the field oxygen transfer coefficients with gas flow for both nonsteady and steady state analyses. The degree of scatter for the steady state analysis (β = 0.95) is significantly greater than that for the nonsteady state results. This may be due to poor probe calibration since the $K_La$ values for the steady state analysis are dependent on probe calibration while for the nonsteady state analyses, the $K_La$ estimate is independent of calibration as long as it is constant over a run. Both analyses exhibit curvature when the line of best fit (by eye) is assumed to pass through the axis.

When the average $K_La$ value from two D.O. probes was used for the dual nonsteady state results, the range of values is presented. The difference between the actual $K_La$ values and the mean is generally less than 6%. For Run #2 a difference of 12% was obtained possibly due to different probe locations near the tank effluent, one at the tank center at a depth of 10 ft. and one at the tank wall at a depth of 2 ft. For Run #5 conducted in Grid #2 a marked spatial difference in $K_La$ values existed, with the probe in the center of the segment yielding significantly higher $K_La$ values than the downstream probe. This was probably due to the unequal gas flow distribution existing in the second Grid, the average $K_La_f$ value used to represent the overall grid segment analyzed. The curve through the nonsteady state data yields oxygen transfer coefficients about 7% higher than that through the

## TABLE 2. MEASURED AND ESTIMATED GAS FLOWS, UPTAKE RATE AND SATURATION VALUES FOR WHITTIER NARROWS RUNS

| | MEASURED VALUES | | | ESTIMATED WASTEWATER DISSOLVED OXYGEN SATURATION CONCENTRATION $C^*_{\infty f}$ | |
|---|---|---|---|---|---|
| | Gas Flow, scfm | | | | |
| Run | Low $G_{s1}$ | High $G_{s2}$ | $O_2$ Uptake Rate mg/ℓ-hr | $\beta = 0.95$ | $\beta = 0.99$ |
| 1 | 520 | 960 | 35.4 | 8.2 | 8.5 |
| 2 | 500 | 1190 | 29.8 | 8.2 | 8.5 |
| 3 | 590 | 1160 | 34.2 | 8.1 | 8.4 |
| 4 | 690 | 1230 | 30.8 | 8.1 | 8.4 |
| 5 | 390 | 1440 | 22.0 | 8.1 | 8.4 |
| 6 | 330 | 1160 | 18.5 | 8.2 | 8.5 |
| 7 | 600 | 1130 | 31.0 | 8.2 | 8.5 |

## TABLE 3. DUAL NONSTEADY STATE RESULTS FOR WHITTIER NARROWS RUNS

| | | | Steady State $O_2$ Concentrations, mg/ℓ | | Calculated | |
|---|---|---|---|---|---|---|
| Run | $K_La_{f1}$ | $K_La_{f2}$ | $C_{R1\infty}$ | $C_{R2\infty}$ | $O_2$ Uptake Rate, R, mg/ℓ-hr | $O_2$ Saturation Value, $C^*_{\infty f}$, mg/ℓ |
| 1 | 3.7 | 6.1 | 0.7 | 4.1 | 32.0 | 9.3 |
| 2 | - | 6.8 | < 0 | 3.9 | - | - |
| 3 | - | 6.8 | < 0 | 2.9 | - | - |
| 4 | 5.4 | 7.2 | 1.2 | 3.5 | 49.7 | 10.4 |
| 5 | 3.0 | 8.1 | 0.7 | 5.2 | 21.4 | 7.8 |
| 6 | 3.0 | 7.2 | 0.8 | 5.4 | 23.7 | 8.7 |
| 7 | 4.0 | 7.7 | 0.5* | 3.8 | 27.5 | 7.4 |

* Estimated at 0.2 mg/ℓ lower than measured value at last downstream station due to spatial $O_2$ gradient.

TABLE 4. STEADY STATE RESULTS FOR WHITTIER NARROWS RUNS

| | $O_2$ Transfer Coefficients*, $hr^{-1}$ | | | |
|---|---|---|---|---|
| | $\beta = 0.95$ | | $\beta = 0.99$ | |
| Run | $K_La_{f1}$ | $K_La_{f2}$ | $K_La_{f1}$ | $K_La_{f2}$ |
| 1 | 4.7 | 8.6 | 4.5 | 8.0 |
| 2 | - | 6.9 | - | 6.5 |
| 3 | - | 6.6 | - | - |
| 4 | 4.5 | 6.7 | 4.3 | 6.2 |
| 5 | 3.0 | 7.6 | 2.9 | 6.9 |
| 6 | 2.5 | 6.6 | 2.4 | 6.0 |
| 7 | 4.0 | 7.0 | 3.9 | 6.6 |

* $K_La_f = R/(C^*_{\infty f} - C_{R\infty})$

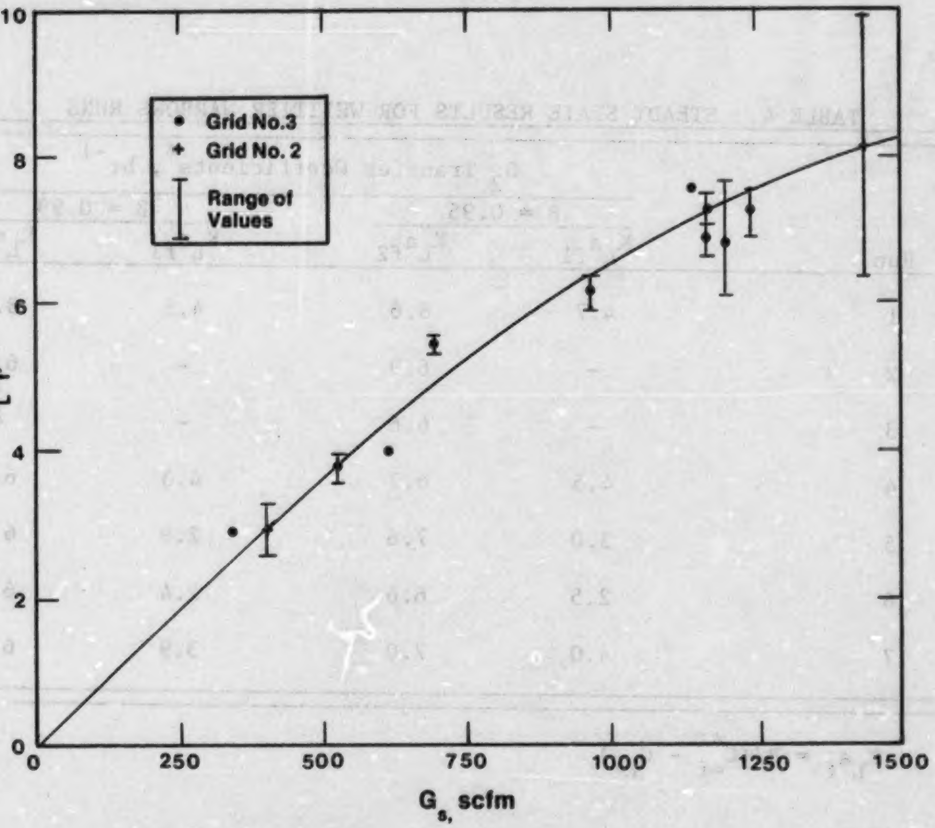


Figure 5. Effect of Gas Flow on $K_La_f$ for Dual Nonsteady State Analysis

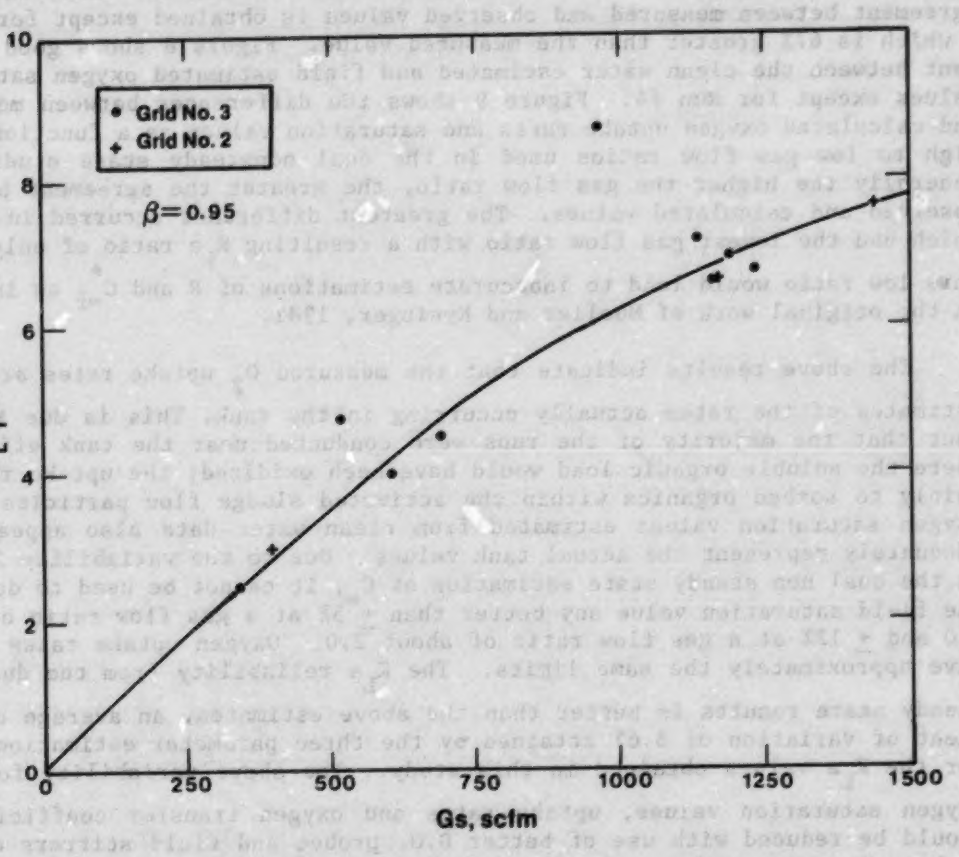


Figure 6. Effect of Gas Flow on $K_La_f$ for Steady State Analysis

steady state data. Since the higher $\beta$ value of 0.99 would yield still lower steady state $K_La$ values, the $\beta$ of 0.95 was used in further comparisons.

The oxygen uptake rates calculated by the dual nonsteady state analysis are compared in Figure 7 to the average measured values for each run. Good agreement between measured and observed values is obtained except for Figure 4 which is 67% greater than the measured value. Figure 8 shows good agreement between the clean water estimated and field estimated oxygen saturation values except for Run #4. Figure 9 shows the differences between measured and calculated oxygen uptake rates and saturation values as a function of the high to low gas flow ratios used in the dual nonsteady state studies. Generally the higher the gas flow ratio, the greater the agreement between observed and calculated values. The greatest difference occurred in Run #4 which had the lowest gas flow ratio with a resulting $K_La$ ratio of only 1.3. This low ratio would lead to inaccurate estimations of R and $C^*_{\infty f}$ as indicated in the original work of Mueller and Rysinger, 1981.

The above results indicate that the measured $O_2$ uptake rates are good estimates of the rates actually occurring in the tank. This is due to the fact that the majority of the runs were conducted near the tank effluent where the soluble organic load would have been oxidized; the uptake rate due mainly to sorbed organics within the activated sludge floc particles. The oxygen saturation values estimated from clean water data also appear to adequately represent the actual tank values. Due to the variability inherent in the dual non steady state estimation of $C_\infty$, it cannot be used to delineate the field saturation value any better than $\pm$ 5% at a gas flow ratio of about 4.0 and $\pm$ 12% at a gas flow ratio of about 2.0. Oxygen uptake rates should have approximately the same limits. The $K_La$ reliability from the dual non steady state results is better than the above estimates, an average coefficient of variation of 3.6% attained by the three parameter estimation model for the $K_La$ values obtained in this study. The above variability for the oxygen saturation values, uptake rates and oxygen transfer coefficients should be reduced with use of better D.O. probes and field stirrers on the probes when the probes require a minimum velocity across the membrane. Figure 10 summarizes the oxygen transfer efficiencies attained for each run as a function of the gas flow rate per diffuser. The oxygen transfer efficiencies (OTE) are calculated at zero dissolved oxygen concentration using the field $K_La$ values and clean water saturation values at 28°C as follows:

$$\frac{OTE}{100} = K_La_f \; C^*_{\infty f} \; 8.34 \; V/1.04 \; G_s$$

where: $K_La_f$ [=] $hr^{-1}$

$C^*_{\infty f}$ [=] mg/ℓ

V = total volume of one Grid, 0.322 MG

$G_s$ [=] scfm

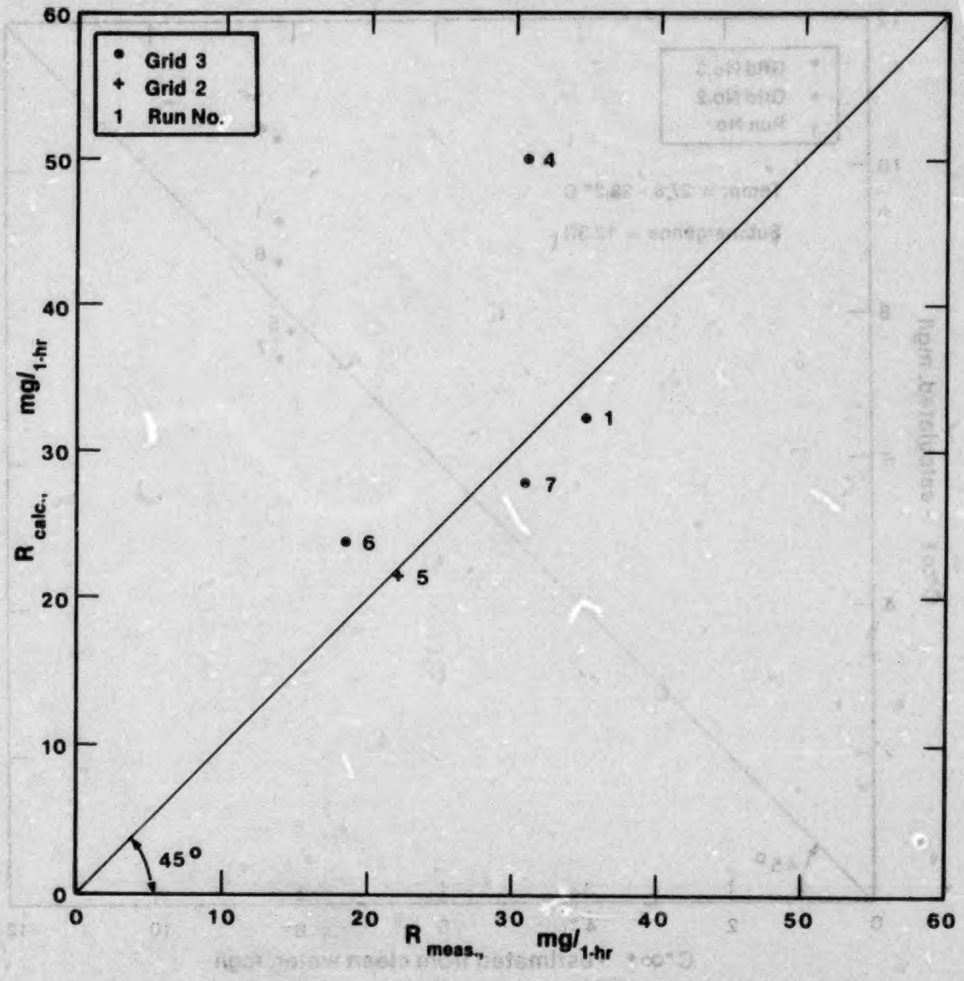


Figure 7. Comparison of measured and calculated $O_2$ uptake rates

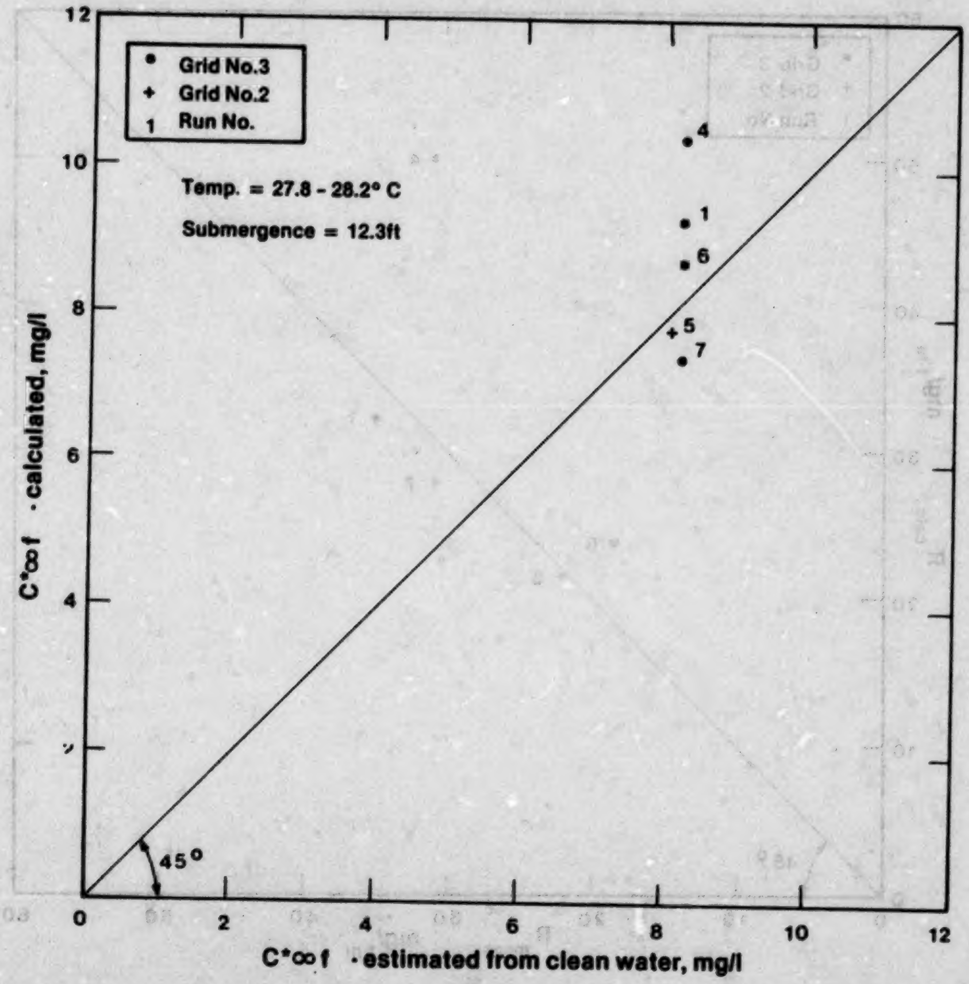


Figure 8. Comparison of Calculated and Estimated (β = 0.95) Oxygen Saturation Values.

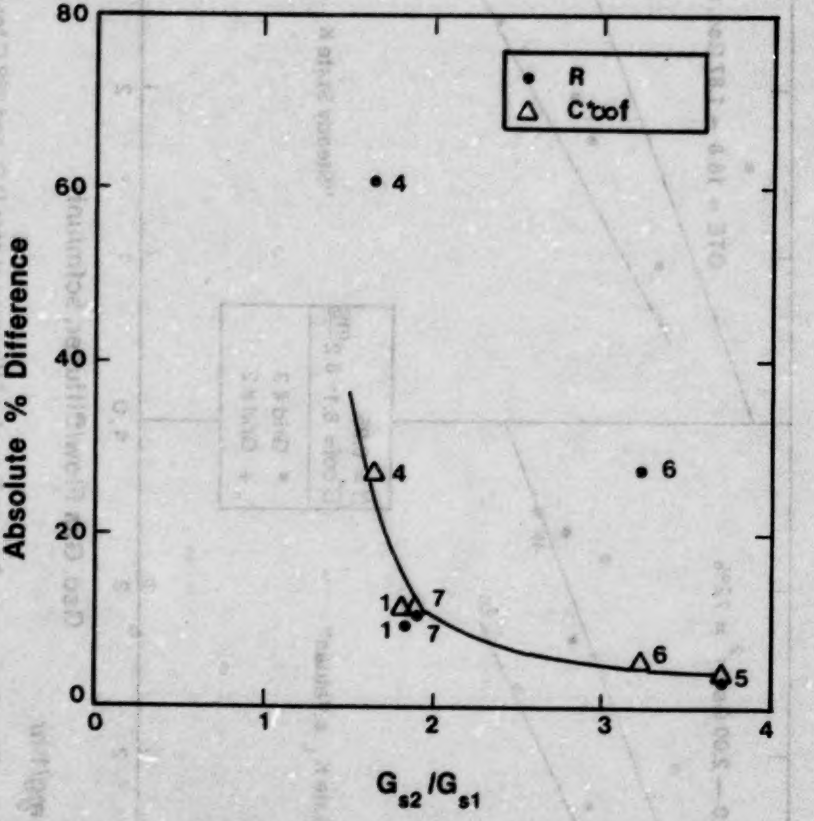


Figure 9. Effect of Gas Flow Ratio on Difference Between Measured and Calculated $O_2$ Uptake Rates and Saturation Values

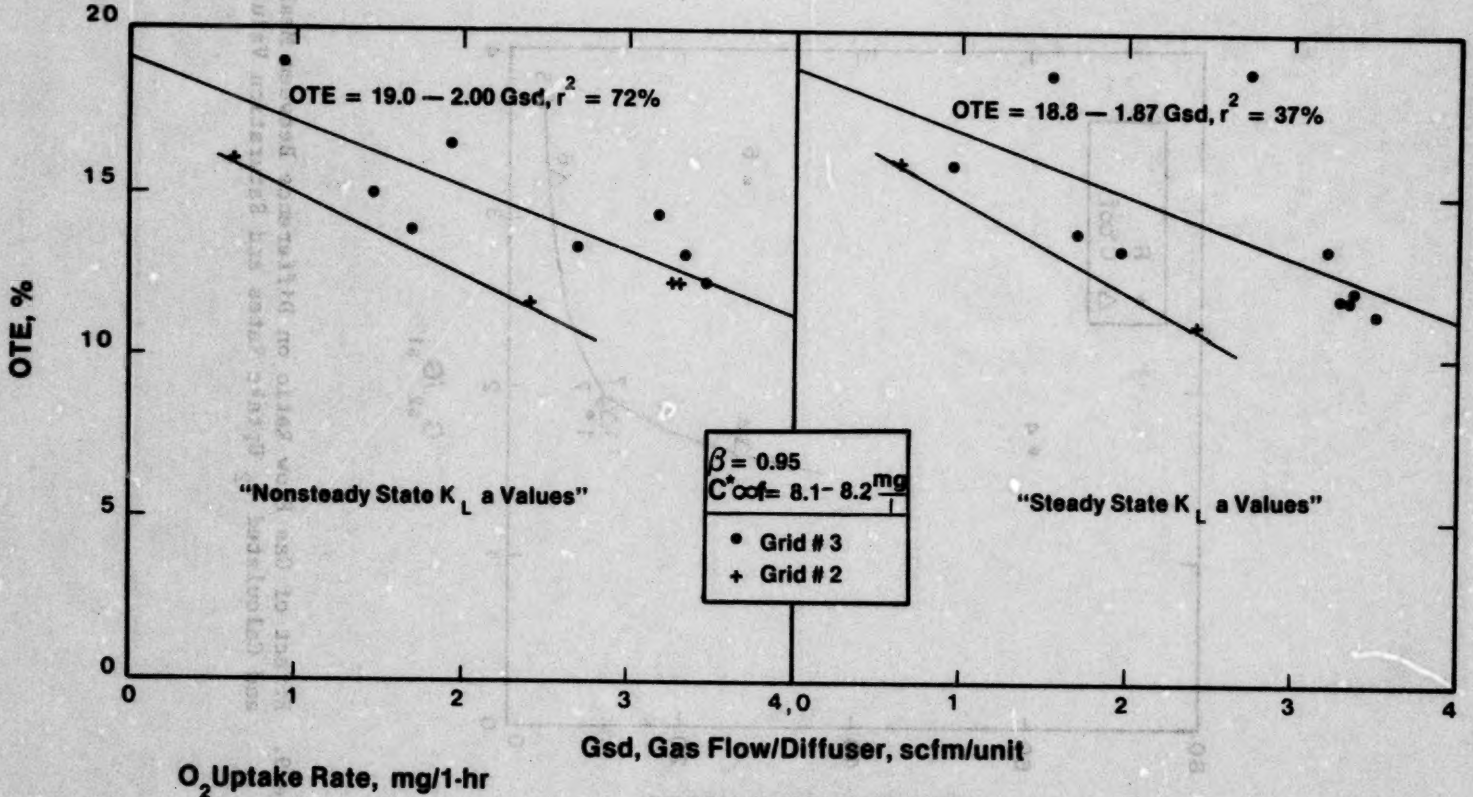


Figure 10. Effect of Gas Flow/Diffuser on Oxygen Transfer Efficiency at zero D.O. and 28° C for Non Steady and Steady State $K_L$ a values

For both Grids 2 and 3, transfer efficiency decreases with increased gas flow/per diffuser. A linear regression on the Grid #3 data gives similar results except a significantly greater correlation coefficient is obtained using the non steady state estimate for $K_La$ over the steady state. For the non steady state estimates only 28% of the variability is due to factors other than gas flow differences while the steady state results show 61% of the variability in OTE is due to factors other than gas flow change. Although the data base is small this does tend to indicate probe calibration deficiencies as discussed previously. The lower OTE values for Grid #2 may be due to possible plugging problems or the greater diffuser density in this grid compared to #3 with similar total gas flows giving similar OTE values. Reduced alpha values in Grid #2 compared to Grid #3 would also provide this type of result, however the Grid #2 data was attained at half flow at relatively low oxygen uptake rates where the alpha value may have been close to that normally present in Grid #3.

CONCLUSIONS

1. Both the dual nonsteady state and steady state analyses are capable of measuring oxygen transfer coefficients under process conditions in long narrow aeration tanks when constant oxygen uptake rates are obtained.

2. For the Whittier Narrows study, the average non steady state $K_La_f$ results showed somewhat less variability than the steady state results. This was caused to some extent by lack of field stirrers on the D.O. probes and possible inaccuracies in probe calibration.

3. In the dual nonsteady state analyses, a gas flow ratio of 2:1 provides estimates of oxygen uptake rate and saturation value within ± 12% of steady state uptakes and saturation values estimated from clean water data. This variability is reduced to ± 5% at a gas flow ratio of 4:1. Unless this variability can be further reduced by using higher gas flow ratios and/or better D.O. probes, clean water saturation values properly corrected for waste total dissolved solids should be used as field saturation values.

4. For the fine bubble diffuser system studied at Whittier Narrows, both nonsteady state and steady state results showed oxygen transfer efficiencies (OTE) to decrease with increasing gas flow per diffuser.

RECOMMENDATIONS

1. For a long narrow (plug flow type) aeration tank, dual non steady state testing should be conducted in a section with uniform uptake rate and $O_2$ transfer rate using two sets of probes, one set near the upstream end of the section to define the boundary effects and the other some minimum distance downstream approximately 2.5 to 3.0 times the ratio of the forward velocity to the oxygen transfer coefficient ($U/K_La_f$). A minimum of two

probes should be used at the downstream station which are calibrated at the wastewater temperature and contain field stirrers when required by the particular probe. Oxygen uptake rates should be run at three locations in the section during the testing to insure minimal temporal and spatial variability. Steady state dissolved oxygen profiles should be obtained in the section before and after each test. Low gas flows ($G_{s1}$) should maintain minimum dissolved oxygen levels of at least 0.5-0.7 mg/ℓ at all points in the section being analyzed. High gas flows ($G_{s2}$) should be set as high as possible to attain maximum accuracy for $O_2$ uptake rate and saturation values. $G_{s2}/G_{s1}$ ratios of at least 1.8/1 should be attained.

2. Steady state testing can be conducted in a tank section similar to that above. Both oxygen uptake rates and steady state dissolved oxygen profiles should be obtained in the section at a period of relatively constant flow and organic loading to the tank. Data analysis should be conducted using the dissolved oxygen concentration at spatial equilibrium, $C_{R\infty}$.

## REFERENCES

Mueller, James A. and Rysinger, J.J. "Diffused Aerator Testing Under Process Conditions," 36th Annual Purdue Industrial Waste Conference, May 1981.

## ACKNOWLEDGEMENTS

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## DISCLAIMER

**The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency, and no official endorsement should be inferred.**

# APPENDIX

## A. DUAL NONSTEADY STATE TESTING PROCEDURE

The following procedure was used during the Whittier Narrows study.

1. Set up probes and recorder on aeration tank

2. Utilize tap water at temperature equilibrium with aeration tank for probe calibration. This was accomplished by submerging a plastic bottle containing tap water in the tank for ∿ ½ hour. Two Winkler tests were conducted and the contents poured into a BOD bottle for probe readings using a magnetic stirrer for agitation. For the larger probes, a bucket at temperature equilibrium was used for calibration.

3. Calibrate recorder to probe readings.

4. Insert probes in proper position in aeration tank.

5. Measure manometer readings across the orifice plate on the downcomer for each grid and tank D.O. values to insure D.O. not limiting. Also measure in line air temperature and pressure.

6. Change valve on main header for gas flow diversion from the two aeration tanks not being studied to the one under study. Open valve on downcomer of grid being analyzed to attain maximum gas flow. This procedure was accomplished in less than one minute.

7. Continually record D.O. readings either manually or with the recorder for the test duration. Test durations were 25 to 40 minutes for the high gas flow and 40 to 60 minutes for the low gas flow.

8. During the test, sample at each probe location for $O_2$ uptake rate. Each uptake rate required approximately 10 minutes, duplicate uptake rates obtained at each location during the total study.

9. When steady state was reached for the high gas flow, the valve changes were reversed.

10. D.O. readings were again recorded and uptake rates measured until steady state was attained.

11. Upon completion of the study, probe calibration was checked using the procedure in #2.

12. Other parameters such as flow rate (raw and return sludge), tank water depth and atmospheric conditions (barometric pressure, temperature and relative humidity) were also obtained.

B. Equations for Dual Nonsteady State Analysis

1. Batch System or Plug Flow System

a. $K_La_f$:

$$C_R - C = (C_R - C_o)e^{-K_La_ft}$$

where: $C_R$ = steady state concentration at a location in the aeration tank a minimum distance of 2.5 to 3 x $U/K_La_f$

$C$ = probe reading at any time t at the above location

$C_o$ = probe reading at zero time at the above location

$K_La_f$ = field oxygen transfer coefficient

$t$ = time

$U$ = longitudinal velocity in aeration tank = $(Q_i + Q_R)/A_c$

$A_c$ = tank cross sectional area

$Q_i$, $Q_R$ = raw and sludge recycle flow rates, respectively.

The three parameter estimation model gives both $K_La_f$, $C_R$, and $C_o$ for each non steady state test.

b. R:

$$R = (C_{R2\infty} - C_{R1\infty})/(K_La_{f1}^{-1} - K_La_{f2}^{-1})$$

where: R = Oxygen Uptake Rate

Subscripts 2 and 1 = high and low gas flow conditions respectively.

Subscript $\infty$ = value at spatial equilibrium

where $\frac{dc}{dx} = 0$

c. $C^*_{\infty f} = C_{R\infty} + \frac{R}{K_L a_f}$

Either condition 1 or 2 may be utilized for this computation. $C^*_{\infty f}$ is assumed constant for both gas flows.

2. Completely Mixed System

a. $K_L a$:

$$C_R - C = (C_R - C_o)\, e^{-(K_L a_f + 1/t_o)t}$$

where: $t_o$ = detention time, $V/(Q_i + Q_R)$

V = volume of completely mixed tank

b. R:

$$R = \frac{C_{R2} - C_{R1} + \frac{1}{t_o}\left(\frac{C_i - C_{R1}}{K_L a_{f1}} - \frac{C_i - C_{R2}}{K_L a_{f2}}\right)}{K_L a_{f1}^{-1} - K_L a_{f2}^{-1}}$$

c. $C^*_{\infty f}$:

$$C^*_{\infty f} = C_R + \frac{R}{K_L a_f} - \frac{C_i - C_R}{K_L a_f\, t_o}$$

Either condition 1 or 2 may be utilized for this computation. $C^*_{\infty f}$ is assumed constant for both gas flows.

3. Nonsteady State Plug Flow Equations at any position in aeration tank:

a: Assumptions:
(1) Constant R
(2) Constant $K_L a_f$

(3) Initial D.O. profile at steady state in aeration tank at any initial $K_L a_f$ value

$$C_{R\infty i} - C(x,o) = (C_{R\infty i} - C_o)e^{-K_i x/U}$$

$C(x,o)$ = initial steady state concentration at any x

$K = K_L a_f$

subscript i = initial condition

$C_o$ = initial D.O. at boundary

(4) Upstream boundary condition can be described by an exponential function similar to nonsteady state aeration equation.

$$C_m - C(o,t) = (C_m - C_o)e^{-K_o t}$$

$C_m$ = steady state concentration at boundary after gas flow change

$K_o$ = coefficient describing rate of change of D.O. concentration at boundary

b. Solution:

(1) For $t - x/U \leq 0$

$$C_{R\infty} - C(x,t) = (C_{R\infty} - C_{R\infty i})e^{-Kt} + (C_{R\infty i} - C(x,o))e^{-(K-Ki)t}$$

where $C(x,t)$ = D.O. concentration at any x at any time $\leq x/U$

$K = K_L a_f$ after change in gas flow

(2) For $t - x/U > 0$

$$C_{R\infty} - C(x,t) = (C_{R\infty} - C_m)e^{-K_o(t - x/U) - K x/U}$$

where $C(x,t)$ = D.O. concentration at any x at any time $> x/U$

**(3) At steady state, $t = \infty$ and Equation (2) becomes:**

$$C_{R\infty} - C(x,t) = (C_{R\infty} - C_m)e^{-K x/U}$$

# NEW DIRECTIONS IN AERATOR EVALUATION

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Aerators must operate within a system composed of tank dimension and particular mixed liquor characteristics. There is a need to be able to evaluate the performance of the entire system, and to be able to sort out the cause of any inefficiencies which may be found.

This paper discusses two approaches designed to evaluate total system performance in terms of aerator performance. Each of these approaches is developing and their limitations are being better definied. Both involve testing respiring systems. The first is the mass balance approach developed to assist in the evaluation of aerated stabilization basins where other tests are impractical. The second is the inert gas tracer method which has the capability to determine the oxygen transfer in almost any system.

## THE MASS BALANCE APPROACH

The mass balance procedure measures the change in the oxidative state of the respiring biological system. The method can be applied to high rate activated sludge systems and low rate aerated stabilization basin systems. The accuracy of the method depends on the extent to which all factors which supply and withdraw oxygen from the system are measured. A number of these factors can be minimized during the test period so that the major factor which determines the aerator oxygen transfer is the reduction in BOD through the respiring system.

The mass balance procedure has been applied to low rate aerated stabilization basin (ASB) systems because oxygen transfer cannot be readily measured using more direct methods. The mass balance method is simple and straightforward especially when the ASB system is aerator (oxygen) limited and where significant settleable biological solids aren't produced (1).

### Minimum DO Procedure

In order to minimize or eliminate a correction factor for the working DO in the basin, it is preferred that the system be oxygen limiting during the test period. Tuning the system may be necessary (i.e. turning off or redeploying aerators) if the system normally carries excess dissolved oxygen. Actually, several tests at approximately 60, 80 and 100% of tuned aeration capacity may be performed to insure an accurate test. If the system is oxygen limited, the oxygen transfer efficiency (kg/kwh) should be equal for each test. If the values don't agree, other limitations may exist, such as insufficient nutrient or the presence of toxic materials.

Certain systems will lend themselves to segregation for the purpose of testing. Tanks in series, or long narrow tanks, would be subject to testing for various groupings of aerators. Thus, it is possible that tests could be made on various portions of the tuned capacity. All portions of each cluster would have to be oxygen limited. The last stages of such a system might be omitted in order to improve the accuracy of the test procedure because these later stages are apt to produce excess DO during a portion of the test period.

The major variables in the calculation are the incoming and exiting mass of BOD, COD, TOC (converted to its oxygen equivalent) or ultimate BOD. In certain cases, such as with many pulp and paper mill works, the $BOD_5$ can be used because the system removes only a small fraction of the COD(1-2). The calculation determines the difference between influent and effluent BOD during the test period. This difference is then divided by the operating power used by the aerators during the same period. The calculation produces the results in terms of the pounds of oxygen transferred per horsepower per day (or kg/kwh).

It should be recognized that under oxygen limited conditions there is a potential to accumulate BOD in the microbial mass which may result in an apparent oxygen transfer rate (OTR) greater than the actual OTR. However, because most ASB systems are respiring at extremely low rates, synthesis effects are minimal and can be neglected.

There are several other adjustments which can be incorporated into the calculation. A discussion of each follows:

1. DO Difference - If the system has been properly tuned, DO returning to the aerator should be zero and DO leaving the basin will probably be zero. Thus, the case reduces to the situation where the incoming wastewater contains DO which would have to be debited from the aerator transfer. For most tuned situations, this factor represents less than 1% of the total transfer.

2. Surface Area - This factor is usually disregarded. However, the aerators could be debited with a value of 50 lbs oxygen/acre of surface. This value represents BOD removal associated with anaerobic stabilization basins which are not mechanically aerated(3). Unless the ASB is very large in comparison to the power combined in the aerators (i.e. greater than $10^6$ gal/HP), the contribution to the surface area to aeration is not a major factor.

3. Benthal Oxygen Demand - Aerated stabilization basins are not completely aerobic. However, the pumping rate of the mechnical or diffused aerators recirculates the total basin contents through the aerobic zone to an extent which impedes the development of highly reduced conditions. Thus, appreciable amounts of oxygen are not required to satisfy an immediate chemical oxygen demand.

However, bottom deposits, no matter how thin, constitute a continued sink for any oxygen which reaches the bottom. Sediment oxygen demand studies on a variety of benthal deposits have shown uptake rates varying from 1.0-20.0 $gm/m^2/day$. A credit of 5.5 $gm/m^2/day$ (50 lbs/acre/day) is suggested for the benthal oxygen demand (4,5,6). Again, even if DO exists near the bottom of most of the basin, the amount of oxygen removed is minimal.

4. Nitrification-Denitrification Factor - The oxygen equivalent of the nitrogen balance will have to be considered in some systems. The aerators should be credited with transferring the oxygen necessary to balance the change in the oxidative state of the nitrogen species entering and exiting the system. However, nitrification is suppressed under oxygen limited conditions (e.g. DO less than 1.0 mg/l) and also ceases at BOD:N ratios of 16 or more(7). Thus, nitrification and denitrification aren't significant in the oxygen limited equilibrium test case and the high BOD:N case.

5. Correcting for Temperature - One advantage of using the mass balance approach is that no temperature correction is required. Biological kinetics call for a doubling of the transfer rate coefficient each 10°C rise in temperature, (i.e. $\theta = 1.072$). (8-10) Thus, an oxygen limited system operating at 10°C would simply require less oxygen (HP) than when operated at 20°C. The oxygen demand would be matched only to the extent that there was aeration capacity in the system.

The transfer rate coefficient, $K_La$, is temperature dependent. The $\theta$ value most commonly used is 1.024. This $\theta$ value offsets the temperature coefficient for the saturation concentration of DO in water. As an example, when the DO is limited (i.e. DO=0) the deficit equals the saturation concentration. The result is that the same mass rate of oxygen is transferred into water at 10°C as at 20°C or 30°C under the oxygen limited condition.

## Excess DO Procedure

Some ASB systems can't be turned down to minimize DO throughout the basin. Several additional factors must be considered in order to compute the oxygen transferred by the aeration equipment when excess DO is present in the basin. These factors are briefly reviewed as follows:

1. Driving Force Factor - If the basin is tested when excess DO is present, the aeration equipment isn't operating at minimum efficiency. Thus, a mass balance based on BOD won't account for the potential transfer which would occur had the test been performed at zero DO. In order to compute the potential transfer, it is necessary to determine beta and the average DO in the basin. The average DO, $\bar{C}_R$, can be determined by testing representative volumes of the basin during the test period. The beta will have to be measured using procedures described elsewhere.

The potential BOD removal can then be increased by the ratio of the oxygen solubility to the oxygen deficit which exists during the test. In order to calculate the credit, a sampling program will have to be developed based on the temporal variation in DO at various points in the basin. Daily sampling at key locations in the basin may be required, especially if BOD load to the basin fluctuates markedly. In systems where the photosynthetic production of DO is a factor, the mass balance may have to be determined on an hourly basis in order to properly proportion the excess DO to the aeration system.

2. Temperature Correction - As mentioned under the minimum DO prodecure, it is $K_La$ which is oxygen dependent. Because $K_La$ isn't a factor in the mass balance calculation, a temperature correction isn't required. The effect of temperature will be included in the average DO in the basin. Again, this is offset in the calculation by the change in saturation provided 0 = 1.024.

3. Nitrification-Denitrification Factor - Because excess DO is present in the system, nitrification is expected to proceed in systems treating significant quantities of nitrogen. In activated sludge systems treating sewage, nitrogen oxidation may account for 5-25% of the total oxygen demand. Thus, it is reasonable to expect that the change in the nitrogen oxygen demand (NOD) across the long-term system may be significant where BOD:N ratios are less than 16:1 in the influent. In this case, this balance will have to be computed in terms of oxygen equivalents.

## Limitations of the Mass Balance Approach

Additional attempts at using the approach have also been somewhat successful in activated sludge plants. More trials which incorporate the method will undoubtedly produce more knowledge of its limitations. The following represent some of the circumstances which may limit the application of the procedure.

First, the test period should be long enough to reduce effects of short-term fluctuation in load, performance, and weather. This is especially important in long term systems. In order to overcome this limitation, it is recommended that a period equal to three theoretical detention times, (V/Q), be used for averaging the data.

Second, the procedure assumes that the system is at equilibrium during the test period. Thus, the system should not be accumulating or releasing BOD which cannot be readily accounted for. The loss of settleable solids may be a sink for BOD in certain systems. Thus, the method is limited to systems where the mixed liquor contents contain less than 0.1 ml/hr settleable solids throughout the test period, or, where the oxygen sink associated with settleable solids can be computed.

The last limitation presented pertains to the fact that the mass balance method hasn't been employed in these cases where the lagoon is photosynthetically active.

## THE INERT GAS TRACER METHODS

The inert tracer methods of oxygen transfer measurement allows the in situ measurement of operating wastewater treatment system oxygen transfer equipment without interference from hydrodynamic, biological or chemical processes. The method is an outgrowth of the laboratory and field investigations of the inert gas tracer method for stream reaeration measurement by Tsivoglou, et al. (11,12,13). The technique relies on three separate tracers, an inert gas tracer to measure gas transfer, an inert dissolved tracer to measure dispersion, and a fluorescent dye tracer to track the other two tracers. Because the gas tracer is inert, the technique may be applied to biologically and/or chemically active systems, like respiring wastewater treatment plants. The use of a dispersion tracer allows gas transfer measurements to be made under the full range of hydraulic conditions, from completely mixed to plug flow. The inert tracer method is applicable to all forms of oxygen transfer equipment, coarse and fine bubble diffused air, turbines and mechanical surface aeration.

Because of its expense, the inert tracer method when fully developed is currently being viewed as a referee technique for other, less costly methods. Current research by NCASI and the Amherst College Department of Physics is directed at (a) reducing the cost of the inert tracer method, (b) determining the precision and accuracy of the technique in a variety of applications and (c) developing the dosing, sampling and analytical protocols needed to carry out the test. The inert tracer method is based on several theoretical considerations in turbulent gas transfer between air and water. A common methematical expression used to described gas transfer in wastewater systems is:

$$\frac{d(\beta C_s - C(t))}{dt} = \alpha k_L a(\beta C_s - C(t)) + r \qquad (1)$$

$C_s$ = dissolved gas saturation concentration for pure water;

$C(t)$ = dissolved gas concentration in wastewater at time, t;

$k_L a$ = gas transfer coefficient in pure water;

$\beta$ = ratio of gas saturation concentration for wastewater to that for pure water;

$\alpha$ = ratio of gas transfer coefficient for wastewater to that for pure water;

$r$ = rate of gas utilization by the system.

Equation 1 is applicable to all sparingly soluble gases. It is also applicable to transfer both into the water ($C(t) < C_s$) and out of the water ($C(t) > C_s$).

Based on an extension of Einstein's law for the diffusion of particles within a viscous medium (14) it is possible to demonstrate that the rate of molecular diffusion within a single medium is inversely proportional to the radius of the diffusing molecule (15). Applying the assumption that the transfer of sparingly soluble gases into and out of turbulent water is limited by diffusion in the water phase, it should be possible to relate the relative transfer rates of various gases to their respective molecular radii. This relationship has been made both theoretically and empirically by Tsivoglou, et al. for several molecular and atomic gases, including oxygen (11,12). Additional relationships of this type have been demonstrated for several light hydrocarbon gases and oxygen (16).

Based on the above theoretical considerations it is possible to relate the transfer of an inert gas tracer ($r = 0$ in Eq. 1) to the transfer of oxygen in a respiring wastewater treatment system. By super-saturating the system with the inert gas tracer and observing the decay (transfer out of the water) of that tracer to equilibrium, $\alpha K_L a$ for the tracer gas may be determined. Then by means of the theoretical and empirical relationship between the gas tracer transfer rate and the oxygen transfer rate, $\alpha K_L a$ for oxygen in the system may be calculated. Because a dispersion tracer is also used, the calculated $\alpha K_L a$ may also be corrected for any hydraulic dispersion in the system.

Several gas and dispersion tracer combinations have been proposed and used for inert tracer measurements of gas transfer. The original tracer combination developed by Tsivoglou, et al. (11,12,13) for stream reaeration work was the radiotracer krypton-85 as the gas tracer and tritiated (H-3) water as the dispersion tracer. Both of these tracers may be sensitively measured by liquid scintillation counting (17). This tracer combination has been used in wastewater treatment systems by Neal and Tsivoglou (18) where it was demonstrated that the inert tracer methodology may be applied to both surface aerators and diffused air systems.

The USGS has developed a non-radioactive tracer system for use in stream reaeration studies (19,20). This tracer system used low molecular weight hydrocarbons (either ethylene or propane) as the gas tracer. These hydrocarbons are measured by gas chromatography (21). In this technique the fluorescent dye is also used as the dispersion tracer. There has been concern voiced about the non-conservative nature of the fluorescent dye tracer and the biological activity of the hydrocarbon tracer gases (22). The USGS tracer combination has not yet been applied to wastewater treatment system aeration equipment.

NCASI is currently developing a non-radioactive tracer system for use in streams and wastewater treatment systems. The gas tracer in this system is atmospheric krypton gas. The krypton will be initially measured by isotope dilution mass spectrometry.

The conservative tracer in the NCASI research is lithium chloride which will be measured by atomic absorption spectroscopy. This tracer combination is expected to overcome the limitations of the USGS tracer system, and build upon the established research of Tsivoglou with other forms of krypton gas.

The protocol of Neal and Tsivoglou (18) for the measurement of wastewater treatment system aeration equipment provides a good example of the direct tracer procedures. The treatment system was dosed with a well-mixed volume of krypton-85, tritiated water and fluorescent dye. A bottle containing the tracers was submersed in the treatment system and shattered by means of an electrically actuated, attached blasting cap. Just prior to dosing, samples were collected in the basin to establish background levels for the liquid scintillation counting.

Sampling in the basin was conducted for about one hour after the tracer dosing. In the early treatment system studies, a flow-through fluorometer was used to detect the fluorescent dye tracer. Samples for scintillation counting were collected at the outlet from the fluorometer flow. Later, in the studies, it was discovered that the fluorescent dye tracer was not needed in highly mixed treatment systems. In the later studies, sampling was accomplished with a manual grab sampler.

The gas and dispersion tracers were analyzed in the laboratory using a three channel liquid scintillation counter. Sample counts were standardized against an automatic external standard. For each field sample, two 2 ml replicate counting samples were counted three times apiece, resulting in six replicate counts per field sample.

In order to calculate the gas transfer rate coefficient from the Kr-85 data it was necessary to correct that data for hydraulic dispersion. This was done by using the ratio of the Kr-85 to H-3 concentrations (R) for each station, rather than the Kr-85 concentration alone in the rate coefficient calculations. This effectively corrected the gas tracer data for dispersion. By preparing a semi-log plot of the ratio, R, versus the sampling time, t, the krypton gas transfer rate coefficient, $\alpha K_L a_{Kr}$, is the slope of the resulting line. The equation for this line, which follows directly from the solution of Equation 1 where r = 0, is:

$$\ln R(t) = - K_L a_{Kr} t + \ln R_o \qquad (2)$$

$R(t)$ = Kr-85 to H-3 ratio at time, t;

$R_o$ = Kr-85 to H-3 ratio at time zero;

$K_L a_{Kr}$ = krypton gas transfer coefficient for pure water;

$\alpha$ = ratio of the gas transfer coefficient for wastewater to that for pure water.

The $\alpha K_L a_{Kr}$ arrived at from Equation 2 can then be used to caluculate the $\alpha K_L a_{ox}$ for oxygen by using the gas transfer rate ratios mentioned earlier. Tsivoglou (11,12) has found that the ratio between krypton and oxygen transfer rates is:

$$K_L a_{Kr} / K_L a_{ox} = 0.83 \pm 0.04$$

Based on the protocol outlined above it is therefore possible to estimate the oxygen transfer coefficient in respiring wastewater treatment systems. Similar protocols would be applicable to the other, non-radioactive tracer combinations like the one under development by NCASI. Based on the work by Neal and Tsivoglou (18) it appears that the direct tracer method should serve as an excellent referee technique for testing oxygen transfer measurements of all types of aeration equipment. NCASI's efforts are directed toward reducing costs thus allowing the method to be more universally available and perhaps become the method of choice in testing aeration systems.

## SUMMARY

In an effort to indicate emerging directions for evaluating aeration systems, two methods have been discussed. The first is inexpensive and uses a mass balance in the oxidation state of the system to arrive at mass transfer. This method has been applied to aerated stabilization systems treating paper industry wastewater and is most suitable for longer detention time systems. The second method is an inert gas tracer method which is applicable to almost any system but which may be expensive compared to other methods. The gases used have been radioactive krypton, ethylene and propane. A method which uses stable isotope krypton is currently under development by NCASI working in conjunction with the Physics Department of Amherst College.

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## DISCLAIMER

The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency, and no official endorsement should be inferred.

## NEW DIRECTIONS - OFF GAS METHODS

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The rate of progress in aeration technology that the agenda of this workshop implies is indeed impressive and encouraging. With the projected long term increases in power costs, the urgent need for such progress is apparent.

Some of the areas in which progress is evident, and in which our company has been active are the following:

1.) Development of tests and proceedures for characterization of porous diffusers prior to, and following use, in-situ, as well as in the laboratory.

2.) Development of diffuser specification requirements that will yield improved performance and economy.

3.) Development of non-process interruptive cleaning proceedures for ceramic diffusers.

4.) Development of equipment and techniques sufficiently sensitive to predict the rate of fouling of porous diffusers and permit the application of cleaning cycles at closer to the optimum frequencies.

5.) Development of analytical equipment and techniques directed towards identification and control of causes of porous diffuser fouling.

6.) Development and application of gas phase techniques to evaluate the efficiency of diffused aeration equipment under field conditions: the so called Off-Gas approach.

7.) The assembly of application data for various diffuser systems through dirty water field testing.

This paper is intended to serve primarily as a progress report of our experiences in off gas work since it was reported by Redmon and Boyle (1), at the 1981 Fall meeting of the ASCE Oxygen Transfer Committee. Since their report did not receive wide circulation, it is attached as an appendix to this paper.

We continue to believe the importance of obtaining reliable methods for dirty water testing cannot be exaggerated. The value lies in design as well as system evaluation. Coupled with solid clean water data for the system, a reliable dirty water test provides a direct approach to the estimation of alpha.

Between the fall of last year and the present, the off gas equipment has been employed in six tanks at the Milwaukee Jones Island Plant and three at the Milwaukee's South Shore Treatment Plant. In all of these tanks the ceramic diffuser plates are disposed more or less uniformly and at a ratio of from .2 to .25 of the total tank area. These tests disclosed specific transfer efficiencies more or less comparable with the full floor coverage ceramic data reported earlier by Redmon and Boyle, (Appendix).

In November of 1981, one tank at the wastewater treatment plant at Berlin, Wisconsin, was tested. This plant has Sanitaire full floor coverage ceramic discs at a ratio of diffuser to tank area of about .08. Specific efficiency of this sytem was found to be comparable to the Whittier Narrows plant reported in Appendix 1.

Also in November of 1981, the wastewater treatment plant of a Wisconsin pulp and paper mill, was tested. This plant is equipped with Sanitaire coarse bubble diffusers in a grid configuration. Although the specific efficiency was roughly comparable with the Madison data in Appendix 1, conclusions of system equivalence should not be drawn without taking into account the differences in configuration, air rate, type of waste, etc.

In December, 1981, tests were conducted in parallel with the radio-active tracer method, as described by Neal and Tsivoglou (2), in one of E I DuPont's industrial wastewater treatment facilities. Hugh Campbell has reported this work in somewhat greater detail in Session 3 of this workshop. This plant was originally fitted with Kenics coarse bubble aerators, the helixes and tubes of which had since been discarded. At transfer efficiencies, (OTE), of around 7%, the offgas and radioactive tracer method agreed within 5%, using either the captured gas rate or the measured applied air rate. This and the above experiences, served to enhance our optimism regarding the accuracy and utility of the method.

The fact that correlation between the two methods requires accurate knowledge of air rate suggested an interesting thought.

In general, liquid pahse methods based on respiration rates, tracers, etc., will yield $K_{La}$ without knowledge of air rate. Off gas methods, on the other hand, will yield OTE without knowledge of air rate. In any system under fixed conditions of submergence, temperature, etc., $K_{La}$ is usually approximately linear with air rate, whereas, under similar constraints, OTE is comparatively insensitive to air rate.

Thus, if the object of the test is to predict air requirements for a given oxygen demand, the off gas method provides the advantage of being far less sensitive to errors in measurement of air rate of the tests than the liquid phase methods are, where any lack of precision in air rate measurement results in a corresponding error in prediction.

During the months of May and June, 1982, eleven tests were conducted at the Munincipal Wastewater Treatment Plant in Brandon, Wisconsin. This plant is a small complete mixed, activated sludge plant, fitted with Aerocleve jets.

The objectives of these tests included evaluation of the precision of the method as well as the correlation between it and the continuous steady state respiration rate method. Results in both regards are presently considered to be inconclusive. One unexplained difficulty, not previously encountered, was the consistent disparity between estimated air rate applied and measured air rate collected. Sixteen tests prior to the Brandon work indicated a mean ratio of air supplied to air collected of 1.05 with a standard deviation of approximately .25. The eleven tests at Brandon yielded comparative values of .81 and .15, respectively. Three different gas sampling patterns were employed at Brandon for collection, and the flux obtained by each were surprisingly similar. Applied air was estimated from the PD Blower curve, and errors of the magnitude observed are not to be expected.

The air rate was not the only anomaly observed. Whereas the transfer efficiencies measured in the first eight tests were comparable with those obtained at Whittier Narrows by Redmon and Boyle (1), the last three tests yielded significantly lower efficiencies. It is our present belief that some, if not all of this inconsistency may have resulted from air leaks into the analytical circuit during some of the tests. Leak tests involving passing Nitrogen through the instrument prior to offgas testing are now routinely applied.

On July 21, 1982, parallel tests were run in a walker process plant in Ridgewood, New Jersey, with the steady state test, the dual non-steady state test, and the non-radioactive tracer test. This data has not been fully analyzed, however, there is reason for encouragement in that good agreement appears to exist between the offgas and the dual non-steady state methods. This encouragement is heightened by the fact that the actual transfer rates during the offgas tests were around 3%, which has heretofore been considered outside of the practical range of the method.

Jeff Brochtrup, a graduate student at the University of Wisconsin, has developed a program for theoretically evaluating the precision of the method from assumed estimates of variation of the principal variables. On the basis of our first assumption, the estimates of precision obtained from the program have been vary favorable. We are presently revising our initial estimates of variation based on the field experience to date and plan to report these findings when complete.

The experience with this equipment over the past eight months has resulted in some new conclusions and some modifications to those reported last October.

The theoretical basis for our confidence in the precision of the method has been expanded, but unfortunately, experimental support therefore has not consistently been forthcoming.

This method appears to be applicable to plug flow and non-uniformly loaded and aerated systems with significantly less error source than the liquid phase systems.

In complete mix systems, the errors resulting from non-representative gas sampling do not appear to be as great as originally expected.

The sampling errors that may occur can be reduced, and the facility of the testing improved, by design modifications to the sampling hood.

Further experience with the Leeds and Northrup Oxygen Sensor and one other polarographic unit, (Teledyne Analytical Instruments, Model 320), have given indication that compensation for temperature, pressure, and humidity may best be made through empirically determined relationships using gases similar in composition to offgas and ambient air. To date, we have not developed all the data necessary to define the empirical relationship required for the instruments we have used. We are also quite sure that other types of instruments may employ different empirical relationships.

**The practical method that we have used, and recommend to others, is to manage the reference and offgas streams to obtain as closely as practical, the same velocity, humidity, temperature, and pressure. We believe it is practical to achieve this control within acceptable limits without unduly complicating the equipment.**

**It has also become apparent that different sensors will have different general requirements. For example, some may require air at or near saturated humidity, whereas, others cannot tolerate humid air and resultant - condensation.**

**We believe periodic standardization or qualification at reduced pressures and/or with reference gases of known composition is desirable.**

**There is preliminary indication that the method may be suitable for reliably measuring oxygen transfer rates at levels of less than 5%.**

**We are not, at this time, in a position to specify the optimum sensing device or the ancillary equipment to go with it, but we do believe the system we have used, described by Redmon and Boyle (1), is workable, and can be further developed into a simple, useful tool for diry water testing.**

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# APPENDIX I - PRELIMINARY FINDINGS: OFFGAS ANALYSIS

BY: David Redmon and Wm. C. Boyle

A REPORT TO THE
ASCE OXYGEN TRANSFER STANDARD COMMITTEE
Detroit, Michigan
Presented: October 6, 1981
Revised: October 30, 1981

## Background

The ASCE work toward an Oxygen Transfer Standard has emphasized the need for a simple, reliable dirty water test. Without it, results of the considerable effort in clear water testing lose much of their significance and utility.

There also exists a need for evaluative dirty water procedures that reveal aerator performance for specific locations within an aeration basin and that may be simply applied to tanks that are non-uniformly loaded and/or aerated. Information on alpha and its variability under various conditions would also be an important side benefit of such a method. Hopefully, the method used could be simple and fast and relatively unaffected by transient conditions, as well as non-disruptive of plant operating conditions.

Consideration of error sources and the sensitivity of various proposed dirty water test alternatives indicated that the offgas method was theoretically quite promising, especially for percent oxygen absorptions in excess of five percent in wastewater, providing precise measurement of differences of oxygen and nitrogen plus inert mole fractions between plant air and offgas could be made.

The above factors led to an attempt to develop the offgas method as a means of evaluating dirty water performance of diffused aeration systems.

## State of Development

Two practical problems had to be overcome to make the method feasible. One was the need for a comparatively large area offgas collection device which was light and easy to handle and with which a reasonably representative gas sample could be obtained. A conceptual design of such a collector was presented at last year's meeting in Las Vegas. Subsequently, a prototype, weighing approximately 200 lbs., with a sample area of 8.75 ft. x 2 ft., was built and tested and found to be quite acceptable.

The second problem was to obtain an oxygen sensor which could precisely detect small differences in the partial pressure of oxygen and be adaptable to in situ measurements.

Our investigation of oxygen sensors, although not exhaustive, did consider paramagnetic sensors, gas chromatographs, and polarographic probes. For various reasons the Leeds and Northrup polarographic probe was selected for trial. This probe, due to its design, requires no transport of oxygen across the membrane when in equilibrium with surrounding conditions; therefore, the presence of water droplets or foulants on the membrane, as well as the gas velocity pass the probe do not affect the readings as they can with the Clark type polarographic probes.

A schematic diagram of the offgas analyzer including the in situ oxygen sensor is shown in Figure 1. In general terms, the offgas sample is drawn through a flowmeter in order to determine the rate of offgas collection over the hood. A small portion of this sample is drawn through an analytical circuit in the analyzer which includes the oxygen sensor. Gas temperature and pressure are monitored and controlled to approximately constant values in the analytical stream so that the difference in the partial pressure of oxygen between main air and offgas may be precisely obtained.

## Theory

### Calculations:

The percentage of oxygen absorption may be obtained by determining the mole fraction of water vapor, carbon dioxide, oxygen and inerts (including nitrogen) in both the reference air stream and in the offgas stream. The percent oxygen transfer is determined using the following expression:

$$\text{OTE} = \frac{O_2 \text{ in} - O_2 \text{ out}}{O_2 \text{ in}} = \frac{G_i(MR_{o/i})\ M_o/M_i - G_i(MR_{og/i})\ M_o/M_i}{G_i(MR_{o/i})\ M_o/M_i}$$

in terms of mole fractions:

$$\text{OTE} = \frac{\dfrac{Y_o(R)}{1-Y_o(R) - Y_w(R)} - \dfrac{Y_o(OG)}{1-Y_o(OG) - Y_{co2}(OG) - Y_w(OG)}}{\dfrac{Y_o(R)}{1-Y_o(R) - Y_w(R)}}$$

Derivation of these equations may be found in Appendix I. See the List of Abbreviations and Symbols at the end of this report for definition of the terms used.

In cases where the same conditions of temperature and pressure for both the reference air and offgas are maintained, the mole fraction of oxygen in the reference air times the ratio of the voltage output of the probe for offgas, to the voltage output of the reference stream is approximately equivalent to the mole fraction of oxygen in the offgas stream. In equation form:

$$Y_o(OG) = Y_o(R)\frac{Mv(OG)}{MV(R)}$$

In using the analyzer in the field, reference air is drawn past the oxygen sensor between each offgas measurement; thus one has a reference check on the probe before and after each offgas observation permitting a simple correction for probe drift, if any is observed. This technique reduces the errors in measuring the difference between offgas and reference stream, and makes the reading more precise and accurate.

The linearity of the probe may be checked periodically by drawing the reference air past the sensor under various levels of reduced pressure while keeping the gas temperature constant.

Under these circumstances the partial pressure of oxygen is directly proportional to the total absolute pressure. The criterion that was applied was that the ratio of absolute pressures for the two conditions divided by the ratio of the meter outputs should fall in the range 0.995 to 1.005.

Employing this test for linearity,it was found that the calibration setting for probe output, in millivolts, was not critical, and essentially the same relative change in the voltage output occurred regardless of the setting with reference air at ambient conditions. To obtain the maximum sensitivity, the reference output should be as close to full scale as practical since probe error is reported to be a fixed fraction of the full-scale reading. It should be emphasized that the above procedure requires precise measurement of the difference in the two gas streams and is not necessarily an accurate determination of the absolute value of the partial pressure of oxygen in either stream.

## Sources of Error and Assumptions

As with all models and methods, various assumptions are employed to simplify the analysis. Some of the assumptions that have been employed in this instance are as follows:

1. Nitrogen and inerts (principally Argon) are conservative; that is, that there is no net absorption or desorption of the constituents in question.
2. Negligible de-nitrification is occurring at the test location.
3. There is insignificant leakage of air into the analytical circuit of the analyzer.
4. The airflow rate to the basin is constant during the test.
5. The barometric pressure is constant during the test.
6. The offgas is at 100% relative humidity at the existing mixed liquor temperature.

In addition to whatever errors are induced by the above assumptions, other potential sources of error have been identified, which include:

1. errors associated with the measurement of temperature,
2. errors associated with the measurement of pressure,
3. errors associated with the measurement of humidity in the reference stream,
4. errors associated with the measurement of dissolved oxygen in the mixed liquor,
5. errors associated with the measurement of carbon dioxide,
6. errors associated with the measurement of the rate of offgas collection,
7. errors associated with the lag in temperature compensation where significant temperature differences exist between the two streams,
8. oxygen absorption which could occur due to the presence of mixed liquor solids on the collector's internal surfaces and downstream piping, and
9. errors associated with neglection of surface re-aeration as may occur in uncovered tanks.

Some additional errors which can be compensated for are as follows:

1. D.O. meter drift,
2. changes in mixed liquor D.O., and
3. differences in submergence beneath and outside the collection hood, due to small differences in water level between the sample and the tank.

Experience to date has been that probe drift is minor after the meter has warmed up a sufficient period of time. One might observe a 0.2 Mv drift during a test out of a 160-170 Mv output for reference air. Due to the short time intervals between successive reference checks (typically 5-10 minutes) and the long term measurement of this effect, a linear interpolation for probe drift has been adopted.

Differences in hydrostatic head underneath and outside of the collection hood are at most a few inches of water, thereby resulting in what is considered a very minor error in oxygen absorption.

## Test History and Results

Prior to going to the Whittier-Narrows Plant (LACSD) in the Los Angeles area,two trial runs of the equipment were conducted. These initial runs, which were performed at the Milwaukee Jones Island Plant and the Nine Springs Plant in Madison, Wisconsin, were quite encouraging and provided an opportunity to make appropriate changes in technique and in operation of the analyzer.

During the week of August 9, two full days of offgas analysis were performed at Whittier-Narrows and, subsequently, two additional days were spent at the Madison Nine Springs Plant. Presently, there are no reliable estimates of the precision and accuracy of the method, although it appears that it compares favorably with other methods and may be within acceptable limits.

While at Whittier-Narrows, a complete analysis of tanks 1 and 3 were conducted, as well as some additional data on tank 1 during the dual non-steady state test conducted by Jim Mueller. Tank 1 was installed with a ceramic fine bubble grid system while tank 3 had a jet system. The nozzles were placed along one sidewall and were directed across the tank facing the opposite sidewall. Due to time limitations and operational problems associated with tank 2, which was fitted with porous plastic tubes, no offgas data was obtained for this system.

At the Madison Nine Springs Plant, data on three types of diffused aeration systems were obtained. These systems include ceramic fine bubble grid, a fine bubble tube system installed in a cross-roll pattern, and a coarse bubble spiral-roll system. During this investigation, respiration rate data were obtained in parallel with the offgas method, the results of which will be separately reported.

The preliminary results of the offgas analysis for Whittier-Narrows and Madison are presented in Table 1. It should be emphasized that these findings are reflective of the present state of the art of the offgas method, and may or may not represent the absolute performance of the aeration devices or patterns employed. The relative performance, however, may be of significance.

In order to compare the performance of percent oxygen absorbed of the various systems tests on a more or less equivalent basis, the parameter specific oxygen transfer efficiency, $OTE_{sp20}$, was employed. This value was calculated by dividing the percent oxygen absorption by the estimated effective driving force. These and other miscellaneous calculation procedures appear in Appendix II. At Whittier-Narrows, the dirty water saturation values, $C^*_{ww}$, were estimated by correcting the clearwater values previously obtained for temperature, pressure and air rate, and multiplying by an estimated value of beta; in this instance 0.95. At Madison, they were obtained by similar interpolative corrections to other clearwater data of systems comparable to those tested at Madison. A beta value of 0.95 was assumed in this case also.

Since the clearwater C* value for various diffused aeration systems are different, the specific OTE values are not always directly comparable. The upper portion of Table 1 indicates the $OTE_{sp20}$ for each of the systems tested at Whittier-Narrows and Nine Springs. Also shown is the range and progression of $OTE_{sp}$ inlet to outfall.

Knowing the specific clearwater OTE and the mean weighted average specific oxygen transfer efficiency, $\overline{OTE}_{sp20}$, based on the offgas analysis, permits the computation of alpha. It should be obvious that if the estimate of the clearwater performance is inaccurate that the resulting estimate on alpha will be similarly inaccurate.

As with the specific OTE, Table 1 indicates the mean weighted estimate of alpha, its range, and its progression inlet to outfall. It is interesting to note the values of alpha for the various systems, as well as the relative ratios among the systems tested. It is interesting to note that, in general, alpha seems to rise as the waste approaches the aeration tank outfall. It was observed that a much greater apparent relative change in alpha occurred at Madison as compared with Whittier-Narrows. Much additional work with this technique is required before definitive conclusions regarding the absolute values and changes in alpha with treatment can be established.

The lower portion of Table 1 compares the applied air rate in scfm per square foot of liquid surface to the measured rate of offgas collection. For the most part the measured offgas collection rate agrees quite well with the estimated application rate. Since the offgas rates are used primarily to weight the OTE results and are not needed for mass balance computations, the differences that exist are not considered to be critical.

## Questions and Anomalies

As with most new approaches, various questions and anomalies arise. In an attempt to verify the offgas results, parallel gas chromatograph tests were conducted at the Nine Springs Plant. The agreement was not particularly good, and the reasons for this have not as yet been fully determined. Further parallel tests are planned. In considering all of the details regarding these differences, it was felt that greater confidence should be placed in the probe readings in this case.

An additional question remains regarding the offgas results obtained during the dual non-steady state test conducted at Whittier-Narrows. The test was conducted by observing D.O., respiration rate, etc., at one air rate and then sequentially another at about 2-3 times the magnitude of the first, and then a third test at about the same air rate as the first. Each of these air rates were maintained until equilibrium conditions were obtained.

Lower than expected results were encountered by the offgas method during the period of high airflow. At present, it is not clear why the anomalies in performance were found during this test sequence. They may be related to measurement errors, transient conditions or other unknown causes. Further tests with the offgas method under similar conditions will be undertaken in the future to further explore this apparent discrepancy.

## Tentative Conclusions

1. In the experiences with the offgas equipment to date, the method appears to give reasonable results in most instances.
2. The method appears to be of comparable or better accuracy and precision than other dirty water test methods currently known.
3. The method is simple and fast and relatively inexpensive with little, if any, process disruption.
4. This method is the only one known which measures performance locally in a tank.
5. The method can be used in basins with D.O. levels of less than 0.5 mg/l.
6. The offgas method can yield useful data to determine alpha and its variability throughout the basin under typical operating conditions.
7. A few additional unanswered questions on the offgas method still remain.

8. The data in Table 1 is considered reflective of the present state of the art, and any conclusions drawn from it are considered tentative at best. Substantial additional work is required before firm conclusions should be drawn.

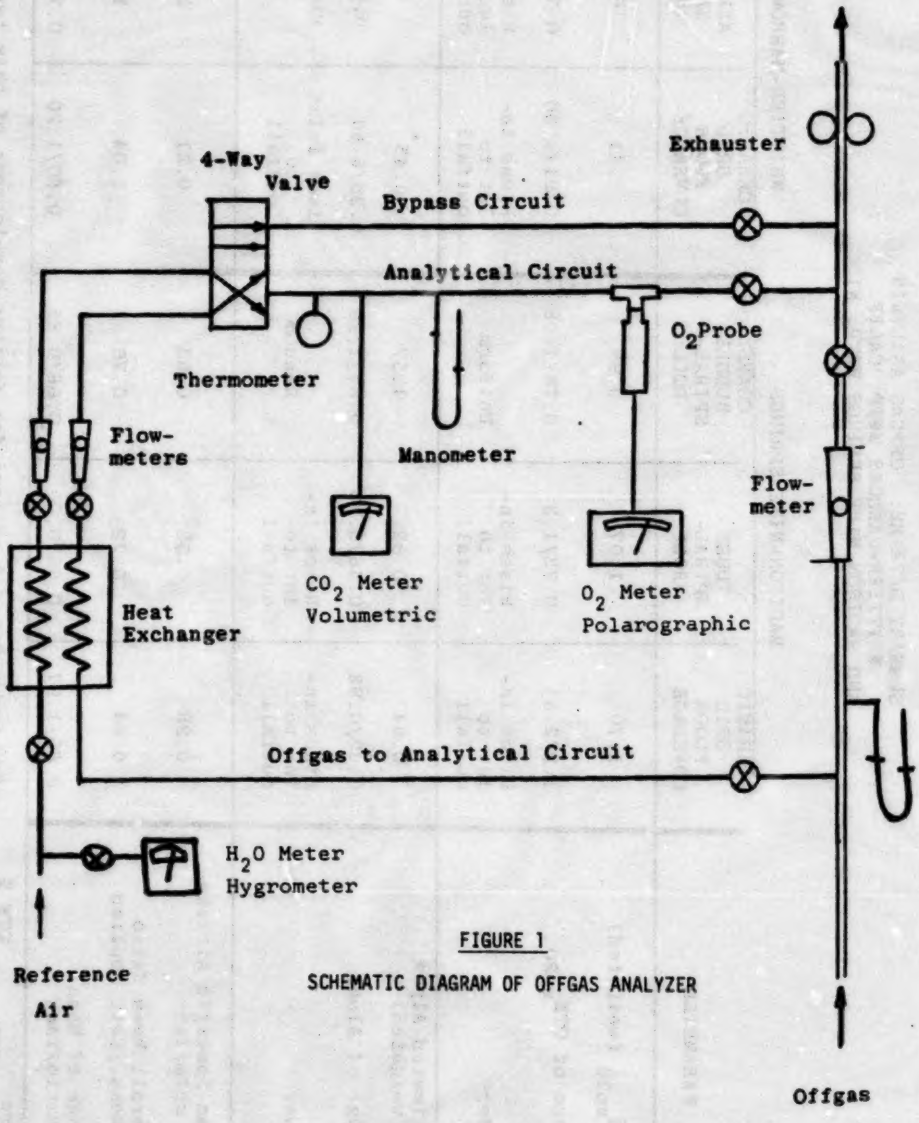
4-Way Valve
Exhauster
Bypass Circuit
Analytical Circuit
O2 Probe
Thermometer
Flow-meters
Manometer
Flow-meter
CO2 Meter Volumetric
O2 Meter Polarographic
Heat Exchanger
Offgas to Analytical Circuit
H2O Meter Hygrometer
Reference Air
Offgas

FIGURE 1

SCHEMATIC DIAGRAM OF OFFGAS ANALYZER

## TABLE 1

### SUMMARY DATA RE: OFFGAS ANALYSIS
### WHITTIER-NARROWS WWTP, CALIF.
### AND MADISON NINE SPRINGS WWTP, WISC.

| PARAMETER | MADISON-NINE SPRINGS | | | WHITTIER-NARROWS | |
|---|---|---|---|---|---|
| | CERAMIC GRID FLOOR COVERAGE | TUBES SPIRAL-ROLL | COARSE BUBBLE SPIRAL-ROLL | CERAMIC GRID FLOOR COVERAGE | JET AERATION SPIRAL-ROLL |
| $\overline{\text{OTE}}_{sp20}$ (weighted) | 1.70 | 1.07 | 0.99 | 1.11 | 0.89 |
| Range of $\text{OTE}_{sp20}$ | 1.2/2.5 | 0.73/1.3 | 0.78/1.08 | 0.91/1.36 | 0.74/1/11 |
| Order | Rises inlet to outfall | Rises inlet to outfall | Uniform | Rises inlet to outfall | Rises inlet to outfall |
| Estimated Alpha (weighted) | 0.64 | 0.62 | 1.07 | 0.45* | 0.58* |
| Range of Alpha | 0.42/0.98 | 0.46/0.85 | 0.83/1.19 | 0.35/0.60 | 0.48/0.72 |
| Order | Rises inlet to outfall | Rises inlet to outfall | Random | Rises inlet to outfall | Rises inlet to outfall |
| Mean Specific Airrate scfm/ft$^2$ | 0.28 | 0.53 | 0.53 | 0.21 | 0.37 |
| Overall Mean Ratio Meas./Est. Applied | 0.84 | 0.72 | 0.76 | 1.04 | 1.08 |
| Range of Mean: Ratio/Tank | 0.66/1.97 | 0.66/0.90 | 0.68/0.86 | 0.80/1.20 | 0.70/1.46 |

NOTE: $\text{OTE}_{sp} = \frac{\text{OTE \%}}{\text{Deficit}}$ See pages 8 and 9 of test for further development of this term.

*Estimated for clean water test.

# LIST OF ABBREVIATIONS & SYMBOLS

| | |
|---|---|
| $C$ | Dissolved oxygen concentration in the liquid phase. |
| $C^*$ | Equilibrium dissolved oxygen concentration in the liquid phase. |
| $C^*_{ww}$ | Calculated value of $C^*$ and equal to ($\beta C^*_{cw}$ $-C$), where $\beta$ is estimated from the TDS of the mixed liquor. $C^*_{ww}$ represents a hypothetical $C^*$ presuming the mixed liquor has no uptake. |

Subscripts of C:

| | |
|---|---|
| cw | Refers to clearwater |
| ww | Refers to wastewater |

| | |
|---|---|
| D.O. | Dissolved oxygen |
| $G_i$ | Mass rate of inerts (includes nitrogen, argon, and remaining trace elements in air). |
| OTE | Percent oxygen absorption or oxygen transfer efficiency. |
| $OTE_{sp}$ | Specific oxygen transfer efficiency (OTE/deficit). |
| $M_o$ | Molecular weight of oxygen, $O_2$, 32.0. |
| $M_i$ | Molecular weight of inerts, 28.1. |
| Mv | Millivolt output of oxygen sensor. |
| $MR_{o/i}$ | Mole ratio of oxygen in reference air to inerts. |
| $MR_{og/i}$ | Mole ratio of oxygen in offgas to inerts. |
| $Y_o$ | Mole fraction of oxygen. |
| $Y_{co_2}$ | Mole fraction of carbon dioxide. |
| $Y_w$ | Mole fraction of water vapor. |

Subscripts of $Y_o$:

| | |
|---|---|
| (R) | Reference air |
| (OG) | Offgas |

## APPENDIX II

**Offgas Calculations**:

Mole Fraction $O_2$ in offgas, $Y_o(OG)$:

$$= \frac{\text{Absol. Cell Press. Refer.}}{\text{Absol. Cell Press. OG}} \times \frac{\text{Absol. Temp. OG}}{\text{Absol. Temp. Refer.}} \times \frac{Mv(OG)}{Mv(R)} \times Y_o(R)$$

Mole Fraction $O_2$ in Reference Air, $Y_o(R)$:

$$= 0.2095(1-Y_{w(R)}) \qquad (1)$$

Mole Ratio of oxygen to inerts in reference or offgas stream:

$$= MR_{o/i} \text{ or } MR_{og/i} = \frac{Y_o}{1-Y_o-Y_{co_2}-Y_w}$$

For the reference stream, all $Y_{o(R)}$ refer to reference air and similarly all $Y_{o(OG)}$ refer to offgas mole fractions for the offgas stream:

$$OTE = \frac{O_2 \text{ in} - O_2 \text{ out}}{O_2 \text{ in}} = \frac{\cancel{G_i(M_o/M_i)}\, MR_{o/i} - \cancel{G_i(M_o/M_i)}\, MR_{og/i}}{\cancel{G_i(M_o/M_i)}\, MR_{o/i}}$$

$$OTE = \frac{MR_{o/i} - MR_{og/i}}{MR_{o/i}}$$

Substituting the detailed expressions for $M_{o/i}$ and $M_{og/i}$ and simplifying yields the following approximate solution:

$$OTE = \frac{\dfrac{Y_{o(R)}}{1-Y_{o(R)}-Y_{w(R)}} - \dfrac{Y_{o(OG)}}{1-Y_{o(OG)}-Y_{co_{2(OG)}}-Y_{w(OG)}}}{\dfrac{Y_{o(R)}}{1-Y_{o(R)}-Y_{w(R)}}}$$

## APPENDIX III

### OTHER MISCELLANEOUS CALCULATIONS

$$OTE_{sp20} = \frac{OTE}{C^*_{ww\infty} - C} \; 1.024^{(20-T)}$$

$C^*_{ww\infty}$ - where clear water value in comparable array at comparable air rate, at standard conditions, $C^*_{20\infty}$ is known:

$$C^*_{ww\infty} = C^*_{20\infty} \frac{P_s + 29.92 \dfrac{C^*_{20\infty} - C^*_b}{C^*_b}}{29.92\left(1 + \dfrac{C^*_{20\infty} - C^*_b}{C^*_b}\right)} \times \frac{C_{sT}}{C^*_b} \times \beta$$

which in most cases may be simplified with acceptable loss in accuracy to:

$$C^*_{20\infty} \frac{P_s \quad C_{sT}}{29.92 \; C^*_b} \beta$$

where $\beta = 1 - .05 \frac{TDS}{5000}$

Where $C^*_{20\infty}$ is not known, but where prior test data has indicated a depth factor as a fraction of submergence that corresponds to the saturation value, $^*d_e$, an estimate of $C^*_{ww\infty}$ may be made:

$$C_{ww\infty} = \frac{[d_e(12/13.5) + P_s]\; C^*_b \times \beta}{29.92}$$

where $P_s$ = station pressure; $C^*_b$ = saturation value book at temperature T; $d_e$ = effective saturation depth at infinite time; $G_o$ = mass ratio of $O_2$ to be applied.

APPENDIX III (continued)

Prediction of performance at similar air rates and the same submergence may be obtained by computing or estimating $C^*_{ww\infty}$ as above and correcting for mixed liquor D.O. and temperature as follows:

$$\text{Oxygen Transfer} = OTE_{sp20}\,(C^*_{ww\infty} - C_F)G_o\theta^{(t-20)}$$

$\alpha$ may be estimated where clear water transfer in similar configuration, air rate, and submergence is known as follows:

$$\alpha = \frac{OTE_{sp20}}{OTE_{Cwsp20}}$$

# FUNDAMENTAL RESEARCH IN BIOLOGICAL WASTE WATER SYSTEMS FOR ADVANCED DESIGN OF MECHANICAL AERATORS

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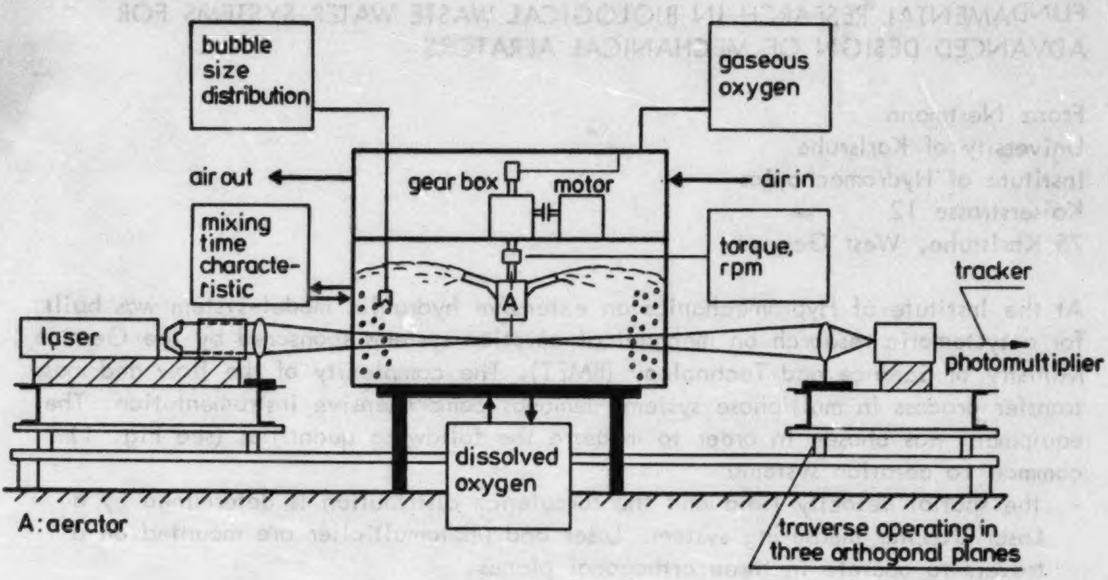
At the Institute of Hydromechanics an extensive hydraulic model-system was built for a systematic research on mechanical aeration systems sponsored by the German Ministry of Science and Technology (BMFT). The complexity of the flow and mass transfer process in multiphase systems demands comprehensive instrumentation. The equipment was chosen in order to measure the following quantities (see Fig. 1) common to aeration systems:

- the spatial velocity field and the turbulence distribution is determined by a Laser Doppler measuring system. Laser and photomultiplier are mounted on a travers to operate in three orthogonal planes.
- the power input is measured by means of the torque on the shaft between aerator and motor.
- the characteristic of the mixing time is detected by tracer.
- the local percentage of dissolved oxygen during the unsteady experiment is measured with electrodes.
- the local bubble size distribution together with the local gaseous air-content are determined in a capillary laser lightened test section.
- the rpm of the 1 kW motor can be selected continuously and is measured by a magnetic field sensor.

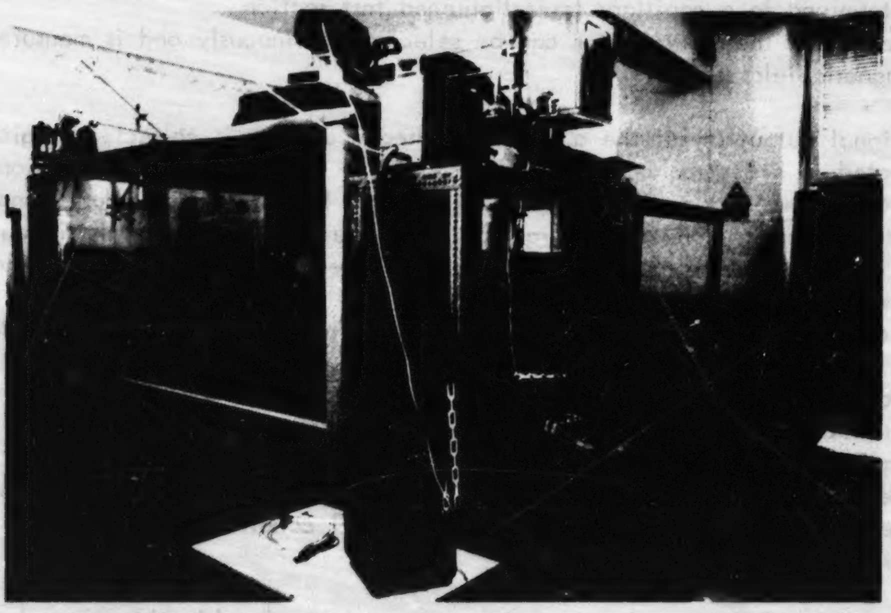
The signal output of all the above mentioned quantities is collected, digitized and evaluated in a desktop computer (1). Up to now the flow field in aeration tanks did not find the adequate attention. Of special interest are boundary conditions like tank geometry and the interaction of the aerator and its geometric environment as well as scale-up effects which are looked into by conducting experiments with models of different sizes. Many of the test series will be compared to results of field experiments later on. Within aeration systems one differentiates between the following single processes:

- local energy input by the aeroator
- interface production and its renewal by the exchange of fluid particles on the liquid side of the interface
- the built-up of concentration gradients $\nabla c$ which is responsible for generating the mass transfer besides the physico-chemical condition and the nature of the interface.

The analysis of the complete aeration process shows that kinetic energy is necessary to induce convective mass transfer. This energy input creates a three-dimensional fully turbulent flow field in the whole basin. In order for the flow to remain fully turbulent it is necessary to overcome the viscous damping by sufficient Reynolds'


bubble size distribution
gaseous oxygen
air out
gear box
motor
air in
mixing time characteristic
torque, rpm
tracker
laser
A
photomultiplier
dissolved oxygen
A: aerator
traverse operating in three orthogonal planes

Instrumentation



View of $5m^3$ basin

Figure 1.

stresses. Production of turbulence is supported by continuous power input into the system. Otherwise the turbulence level subsides due to convection, diffusion and dissipation.

The continuous energy consumption can be described by means of an energy spectrum with the dissipation rate $\varepsilon$ (reduction of kinetic energy per unit mass and time) defined as followed

$$\frac{\partial}{\partial t}\int_0^\infty E(k)dk = \varepsilon$$

With the help of $\varepsilon$ the power input can be determined by

$$P_M = -\varepsilon \quad \text{respectively} \quad P_M = -1/2\,\frac{d}{dt}\left(\overline{u'^2} + \overline{v'^2} + \overline{w'^2}\right)$$

$P_M$ spatially integrated over the basin volume is equal to the power input $P_s$ at the aerotor shaft

$$P_s = 2\pi \cdot N \cdot M_t$$

N: revolution speed 1/s
$M_t$: torque at vertical shaft Nm

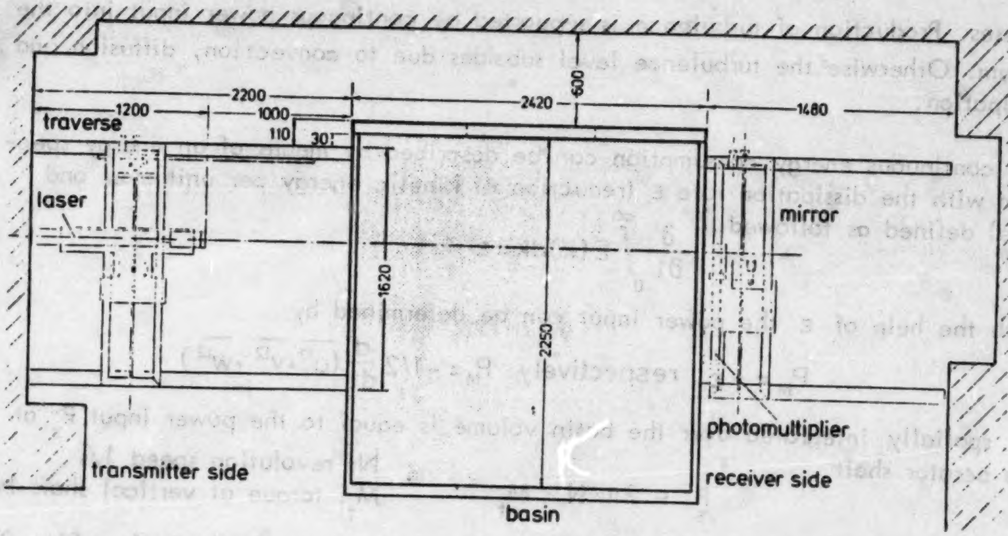
Laser Doppler measurements (2) at different cross sections in the basin (see Fig. 2) give a good inside view of the spatial distribution of the local kinetic energy density. Figure 3 shows the distribution of turbulence intensity v′ for the Karlsruhe Aerotor in vertical direction for a quater basin.

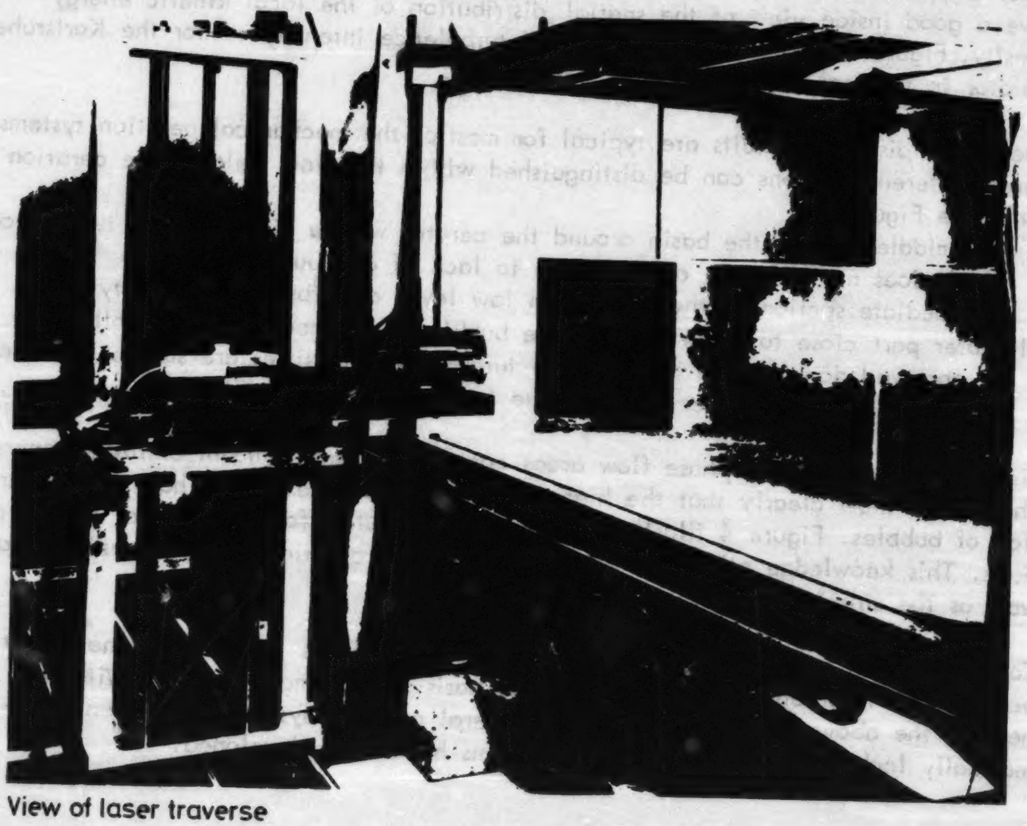
The above discussed results are typical for most of the mechanical aeration systems. Three different sections can be distinguished within the flow field of the aeration tank (see Figure 3):

I The middle part of the basin around the aerator with a high level of turbulence which does not help the aeration due to lack of any bubbles.
II Intermediate section of the basin with low level of turbulence intensity.
III Outer part close to basin walls where bubble entrainment due to a roller can be observed down to the bottom. The turbulence intensities are smaller than in I) because there is higher damping due to the bubble intensity.

Figure 3 shows the multiphase flow areas contain the least amount entrained energy. The results show clearly that the high levels of turbulence do not help the dispersion of bubbles. Figure 4 illustrates the velocity vectors for the vertical cross sections. This knowledge allows to identify the basin areas liable to sedimentation as well as the structure of the macro-vortices.

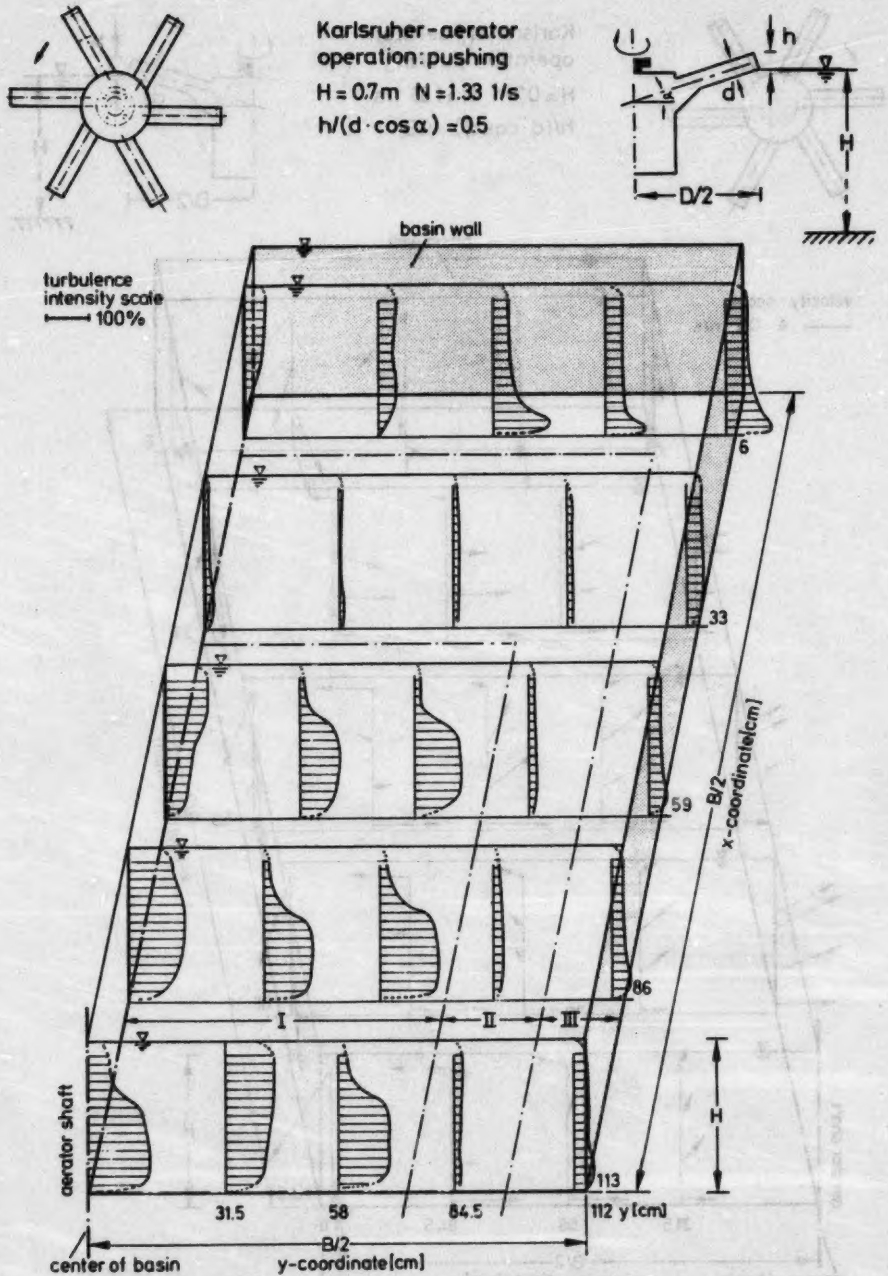
To optimize an aerator system, a detailed research of the flow field is necessary because the turbulent flow structure is the basis for the mass transfer. With the help of the above described techniques several aeration systems have been systematically looked into as well as new systems have been developed.


2200
1200
1000
110
30
2420
600
1480
traverse
laser
mirror
1620
2250
photomultiplier
transmitter side
receiver side
basin



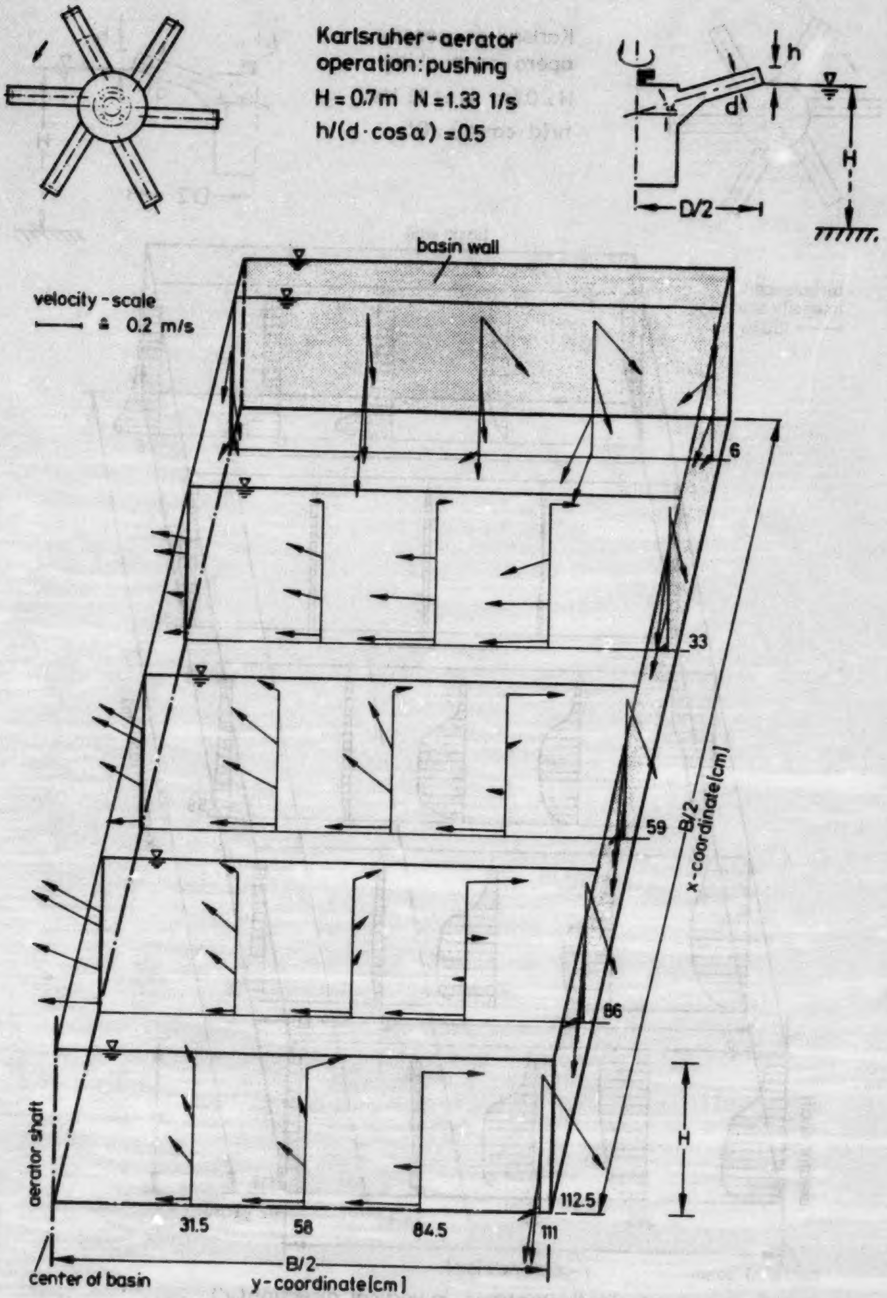
View of laser traverse

Figure 2.



Distribution of velocity fluctuations in vertical direction(v')

Figure 3


Karlsruher-aerator
operation: pushing
H = 0.7m N = 1.33 1/s
h/(d · cos α) = 0.5
h
d
H
D/2
basin wall
velocity-scale
≙ 0.2 m/s
6
33
59
86
112.5
B/2
x-coordinate [cm]
H
aerator shaft
31.5
58
84.5
111
B/2
center of basin
y-coordinate [cm]

Velocity distribution in vertical planes

Figure 4.

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| | | |
|---|---|---|
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| (2) | Lang, C. | Turbulenzmessungen in einer Modellbelebungsanlange mit Hilfe der Laser-Doppler-Anemometrie. Diplomarbeit - Institut fur Hydromechanik - Universitat Karlsruhe (1981). |

## DISCLAIMER

The work described in this paper was not funded by the U.S. Environmental Protection Agency. The contents do not necessarily reflect the views of the Agency, and no official endorsement should be inferred.

# WORKSHOP ON AERATION SYSTEM DESIGN

D. W. Smith

The discussion of the work session on aeration system design centered around three basic themes: the α factor, energy considerations in design, and new aerator system developments. A summary of each theme is presented below.

## Alpha Factor

The factor for adjustment of $K_La$ from clean water to dirty water conditions causes difficulty in making rational design decisions. This is caused by the difficulties in comparing the performance of different types of aeration equipment in dirty water. It is known that the degree of turbulence, the nature of the contaminate and a number of physical components combine together to make prediction and translation of aerator performance information difficult.

The question of how close must an estimate of α be for designs which are to be at capacity in 20 years was passed. The question, although not answered, suggests concern with the level of credibility which must be included in design. It appears that the lack of or limited knowledge of α leads to conservative design decisions so that, to the extent possible, adequate oxygen is transferred to the system. This type of concern leads to a significant amount of effort being spent on specifications and in situ performance testing.

The conditions required for the occurrence of α values greater than one were also reviewed.

## Energy Considerations in Design

The aeration system design must include the aeration chamber if the performance of the equipment is to be maximized. The more complete use of control systems to provide turn down capability and optimum management of the concentration of oxygen in the mixed liquor were noted as needed improvements.

The primary concern is with the total consumption of energy in achieving a desired degree of treatment. Therefore, the control system must ultimately include all components of the treatment facility.

It was noted that the lack of knowledge about the control of unit processes or the lack of ability to control them will impact more as the cost of energy increases. This means the municipalities and industries which require the treatment facilities will have to pay more of their operating funds.

## New Aerator System Developments

Three new approaches to aeration system design were presented. A contractor-activated sludge unit which was designed to use the energy in the air carried into the tank. The system consists of a series of cells between disks on a horizontal shelf. This unit may provide a low total energy cost for the biological treatment step.

The development work on a reverse flow turbine was described. The unit attempts to reduce the amount of energy wasted on limited or no value turbulence. Pilot plant studies on the unit in Germany are about to be started. Model studies are underway.

The third system described was a verticle oxidation ditch consisting of a conventional plug flow tank with a horizontal baffle at mid-depth and a mixing propeller. It was stated that better use of the oxygen added was expected in a prototype unit.

## Participants

D. W. Smith
Sirikun Hopkins
Ray Thacker
Tony Seiler
Michael Marozzi
Frances D. Shepard
Mark Moederer
Steve Black
Henryk Melcer
Mary Anne Lowndes
John Belken
Bill Baumann
Franz Nestmann
Jerry Wren
Robert Alldredge
George Smith
Michael Doyle
James E. Owen

## WORKSHOP ON OPERATION AND MAINTENANCE

D. H. Houck

The operation and maintenance (O&M) workshop convened at 4:30 pm, August 3, 1982, with 20 persons in attendance and D. H. Houck serving as chairman.

Initially, the group discussed the problems of icing and capsizing of floating mechanical aerators in northern climates. It was noted that the spraying effect of these devices serves to cool the aerated liquid to the point of freezing and consequently they cannot be operated during the coldest weeks of the year.

The discussion then turned to fine bubble dome and disc aerators and their O&M problems. Strong sentiments were expressed by the consulting engineers and O&M specialists that the equipment lacks adequate strength in its materials and fabrication, leading to excessive maintenance requirements. Problems cited included failure of plastic parts, notably dome hold down bolts and gaskets, pipe hold down straps, expansion joints, and tank mountings. It was noted that the equipment appears to lack adequate capability to adjust for expansion and contraction brought about by temperature change. The suggestion was made that the design of the equipment follows the English design, where temperature changes are much more moderate. Equipment vendor representatives pointed out that the clients were purchasing the equipment on a low bid basis, causing the level of quality to be driven to the lowest common denominator. It was suggested that engineers need to specify more precisely the degree the materials and "robustness" desired. Also, the manufacturers should be required, and paid for, their direct involvement in the installation of the equipment. If possible, they should assume responsibility for the initial installation as is the English practice. It was noted that the U.K. experience is that failures tend to occur more frequently during the first year after startup, and much less frequently thereafter.

The session adjourned at 5:45 pm.

Participants

Dick Brenner
George Powell
Dave Redmon
Scott Anderson
Rod Gross
Bruce Jank
Chay S. Ching
Jim Laughlin
Tom Pokorsky
Gerry Rupke
Glen Daigger

Read Warriner
Richard Jones
Bill Sturtevant
Tim Kluge
Bob Graf
Forrest Johnson
Paul Cummings
Bob Shevlin
Dan Houck, Chairman

## WORKSHOP ON AERATOR TESTING

W. L. Paulson

This session was chaired by Dr. Robert Baillod, Dr. Hugh Campbell and Dr. Wayne Paulson. Twenty-six people participated in the 75 minute session. The discussion was divided into three parts, 30 minutes for clean water testing, 30 minutes for dirty water testing and 15 minutes for open topic discussion.

## CLEAN WATER TESTING

The clean water discussion focused on the content of the "proposed clean water standard" presented by Baillod and Paulson in Session 3 and included in the printed documents for study at the Workshop.

### Truncation

Two questions related to truncation. One question was "Why truncate the 1st 20% of the DO data and then run all the way to 98% of the DO at saturation?" The 2nd question was "Why deoxygenate all the way to zero DO when truncating anyway?" It was observed that the non-linear method is very sensitive to data closer to the DO at saturation. It was also noted that data early in the run is desirable when evaluating residuals and data adequacy. It has been reported that better quality testing results have been obtained when the DO level in the test tank is depleted to zero for several minutes. Some participants observed that they had conducted successful testing when the DO level only reached 1 or 2 mg/L. The "proposed standard" permits a lower limit of 0.5 mg/L.

### Location of DO Determination Points

The location of DO determination points was discussed. It was noted that there is an emphasis in sensing an equal volume with each determination point. This is recommended due to the averaging of point $K_La$ values in determining the standard oxygen transfer rate. There are spacing limitations cited in the "standard" as well. It was noted that additional DO determination points may be located in corners, near the floor of the tank and elsewhere for evaluating what is occurring in those locations. These points would likely not be part of the oxygen transfer rate analysis. It was observed that if a given system has unusual mixing patterns, equipment placement or geometry the number of DO determination points for the transfer rate analysis should be increased.

### Data Analysis

The following question was asked regarding data analysis. "What about averaging the DO concentration from the several determination points at a given time and make one calculation involving $K_La$ and the oxygen saturation value." It was observed that this procedure would probably be correct.

Based on the experience of several participants, good data yields the same result. It was noted, however, that point $K_La$ data is needed in evaluating other aspects of the testing performance.

Some concern was expressed regarding the calculation of the DO at saturation versus measuring the value at the end of a test. The measured value approach is common to many of the participants. It was observed that the recommended non-linear analytical method uses a calculated value and that it may vary between DO determination points.

### Cobalt Source

It was recommended that consideration be given to changing the stated requirement of reagent grade chemical to permit the usage of a technical grade. This was requested due to the high costs incurred in large field tests or extended testing programs where reagent grade cobalt is used. It was reported that equivalent catalytic action has been observed when comparing the two types. This change in wording will be considered in the revision of the "proposed standard".

## DIRTY WATER

### Oxygen Uptake Measurement

It was reported that there has been as much as a 20% variation in oxygen uptake measurements when comparing the manometric versus the BOD bottle method. There was considerable discussion in an attempt to analyze this discrepancy. The significance is the impact on the dirty water procedure that utilizes the oxygen uptake rate as part of the analysis for oxygen transfer performance. This potential variation was noted with no resolution of the cause.

### Off-Gas Technique

A question was asked regarding the application of this method to plug-flow tanks, eg. What DO concentration do you use? Mr. Lloyd Ewing reviewed the development and application of the method to date in the ASCE study. He also observed that one can use a weighting method where each sample point would have a specific oxygen transfer rate, eg. oxygen transfer efficiency/unit of driving force.

### Hydrogen Peroxide Method

Several participants, including Dr. Kayser, shared their experience in utilizing this procedure. In general, there appears to be a positive reaction to its application for all activated sludge systems. It was observed that the method does not create new errors when contrasted to tracer techniques.

### Tracer Methods

Several participants discussed the methodology and chemistry of the krypton and tritium tracer technique. Dr. Popel reported on the use of helium in the Netherlands.

### Evaluation of Existing Methods

The writer asked the opinion of several participants who are active in dirty water testing for their evaluation of the present methods. They indicated that in their opinion the top four procedures are the hydrogen peroxide method, the radioactive tracer method, the off-gas method for diffused air and the respiration rate approach.

## OPEN TOPICS

### Alpha Analysis

Several participants participated in a review and discussion of the current status of this area of study. Dr. Boyle reported that at this time the ASCE study is comfortable with the use of tall columns for predicting reasonable full-scale alpha values for uniform density fine bubble diffusers. Dr. Boyle and Dr. Kayser indicated that for other systems it is necessary to develop a historical baseline of alpha values with full-scale testing on generic systems in different types of wastewater conditions.

### Detergent Method

The inclusion of this method in the "proposed standard" and its interpretation was discussed. It was noted that the usage of detergent testing in the USA is very controversial. Experience with this approach is quite variable. It was observed that the procedure presented in the "proposed standard" was developed from the method utilized by Dr. Boon and others in the United Kingdom.

Dr. Boon commented on the success of the approach in the United Kingdom where it is utilized to simulate the approximate transfer performance one might expect with domestic wastewater. He reported that in compliance testing one test is conducted with detergent added to clean water in addition to the clean water testing requirement. Dr. Goronszy reported that detergent testing is also conducted in Australia.

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## WORKSHOP ON OXYGEN TRANSFER IN STREAMS

James J. McKeown

The workgroup convened in the afternoon with Messrs. John Sulliven of Wisconsin DNR, Edward Barnhardt of ELBA, Inc. and James J. McKeown of NCASI in attendance. The meeting lasted approximately two hours.

The workgroup undertook a review of the applications where enhancement of the oxygen concentration of receiving waters had been studied. During the discussion twenty-five such sites were identified (Attached A). The workgroup also listed the processes by which DO enhancement had been achieved (Attached B) and discussed which processes were deployed at which sites (Attached A). The workgroup expects to seek additional experience on this topic as well or validate the information assembled on the attachments.

The workgroup also developed a list of factors which have played a role in limiting the utilization of DO enhancement in streams (Attached C). Also, the workgroup knew of two recent situations where the use of oxygen enhancement processes were permitted for meeting DO criteria during critical periods in the year (Attached C).

Finally, work was started on listing some of the technical needs which the group believed necessary to advance the state of the art in applying DO enhancement processes to streams (Attached D). Although the methodology has been in the development stages for decades, it is only now receiving more consideration as a viable alternative to achieving water quality standards. In many respects, its resurgence is a function of a reluctance on the part of many communities and industries to install expensive advanced waste treatment facilities which will only be needed infrequently.

The workgroup also wanted to encourage readers to add to the list of technical needs based on their own experience and perspectives. As a beginning, readers are encouraged to forward such information to:

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(ATTACHMENT A)

## ENHANCEMENT OF OXYGEN IN SURFACE WATERS

### LIST OF RIVERS AND PROCESSES

| | | |
|---|---|---|
| 1. | Black R. SC | Surface Aeration |
| 2. | Millers R. MA | Surface Aeration, Oxygen Diffusion |
| 3. | Miami R. OH | Surface Aeration |
| 4. | Andoscoggin R. ME | Surface Aeration, Downflow Bubble |
| 5. | Willamette R. OR | |
| 6. | Upper Passaic R. NJ | Surface Aeration, Diffused Aeration |
| 7. | Jackson R. VA | |
| 8. | Chicago Canal IL | Diffused Aeration |
| 9. | Trinity R. TX | |
| 10. | Mohawk R. Canal NY | Cascade |
| 11. | Naheola R. AL | Side-Stream Oxygenation |
| 12. | Pearl R. LA | Side-Stream Oxygenation, Turbine Venting, Oxygen Diffusion |
| 13. | Fox R. WI | Turbine Venting |
| 14. | Wisconsin R. WI | Turbine Venting with Oxygen |
| 15. | Peripheral Canal CA | |
| 16. | Thames R. UK | Barge Oxygenation |
| 17. | West Germany | |
| 18. | Red Deer River, Alberta | U-Tube |
| 19. | Parish R. LA | |
| 20. | Neuse R. NC | Side-Stream Oxygenation |
| 21. | TVA Canals | |
| 22. | Flambeau R. WI | Diffused Aeration, Turbine Venting |
| 23. | Las Colinas R. TX | Surface Aeration |
| 24. | Nashua R. MA | Surface Aeration |
| 25. | Penobscot R. ME | Turbine Venting |

(ATTACHMENT B)

## METHODS AND PROCESSES FOR DISSOLVING OXYGEN IN SURFACE WATERS

### Methods

Cascade
Bubble Diffusion, fine and coarse, air or oxygen
Mechanical Surface Aeration
Deep tube or U-tube bubble Contactors
Pressurization
Aspiration

### Processes

Dam Reaeration
In-Stream Aeration, Oxygenation
Side-Stream Oxygenation
Turbine Venting

(ATTACHMENT C)

## LIMITATIONS TO APPLYING PROCESSES FOR DIRECT OXYGEN ENHANCEMENT OF RIVERS

1. Legal or Regulatory
    - prohibitions in specific states or regulations
    - Red Book (i.e. 110% saturation)
2. Aesthetics
    - foam
    - slime
    - noise
    - undersightliness
3. Navigational
4. Engineering
    - protection at high flow for permanent installations
5. Ecological
    - increased scouring or coagulation
    - alteration of biological habitats
    - promotion of slime

Permitted Applications

1. Chicago
2. American Can, Naheola

(ATTACHMENT D)

## TECHNICAL NEEDS

1. Oxygen transfer coefficients of natural water.
2. Knowledge of methods to efficiently mix liquids under high pressure with water at atmospheric pressure without losing the oxygen dissolved at high pressure.
3. Optimize type and placement of sparge ring in turbine venting.
4. Information on the efficiency of introducing fine versus coarse bubbles into the U-tube or downflow bubble contactor.
5. Evaluation of embolysm impact on fish at various dissolved gas levels.
6. Demonstration of the cost effectiveness of the methods based on frequency of occurrence for selected rivers.
7. Investigate the procedure to allocate aeration costs among dischargers.
8. Document costs for Oxygen Production by PSA (est. \$80/ton), cryogenic (est. \$40/ton) and membrane separation (est. ?\$/ton) methods.

## APPENDIX A - SEMINAR/WORKSHOP PROGRAM

## AERATION SYSTEM DESIGN, TESTING, OPERATION, AND CONTROL

August 2-4, 1982

August 2, 1984
0800 Registration - The Wisconsin Center/Lakeshore Room
0850 Workshop Objectives
    W. C. Boyle
    University of Wisconsin
    Madison, Wisconsin
SESSION 1 Chairman - W. C. Boyle
0900 Introduction to Workshop - E. L. Barnhart
0945 Coffee
1000 Types and Characteristics of Aeration Systems - Panel
    Diffused Air - J. Wren
    Mechanical Systems - T. Rooke
    Oxidation Ditch Systems - G. Smith
1045 Translation of Clean Water to Dirty Water - W. C. Boyle and Michael Doyle
1115 System Scale Up - N. W. Schmidtke
1145 Discussion
1230 Lunch
SESSION 2: Chairman - N. W. Schmidtke
1330 Design Protocols - A Global Perspective
    North America - D. Smith
    United Kingdom - A. Boon
    W. Germany - R. Kayser
    Netherlands - H. J. Popel
1215 Discussion
1245 Lunch
1400 Workshop Sessions
    Design - D. Smith (Chairman)
    Operation/Maintenance - D. Houck (Chairman)
    Control - J. Stephenson (Chairman)
    Testing - W. Paulson and C. R. Baillod (Chairman)
1730 Adjourn
OPTIONAL TENNIS TOURNEY

August 4, 1982
SESSION 4: Chairman - W. C. Boyle
0800 New Directions
    J. A. Mueller
    E. L. Barnhart
    J. McKeown
    L. Ewing
0900 Coffee
0915 Summary of Workshop Sessions
1015 Summary of Impressions: Design - Panel
    A. Boon
    R. Kayser
    D. Smith

1115 Discussion
1200 Wrap-Up - E. L. Barnhart
1230 Closure
Afternoon - Optional field trip to Madison Nine Springs wastewater plant to observe off-gas transfer test.
Subcommittee on Oxygen Transfer Standards meeting (open to conferees).

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